Process Plant Design and Simulation

handbook

This book shall serve as a bridge between university and industries It's an honest attempt to make engineering student and young chemical engineers "Ready to use Product" for the industries, so that they don't have to spend 6-month time training new entrants, instead they can work on any real project problem.

The best way to learn process engineering is through solving the real world problems. Simulation software like Aspen HYSYS and FluidFlow etc. are the powerful tools to carry out plant design. And since it has been used by all the design companies, it makes it mandatory for every chemical engineer to learn the same.

With the help of this book, reader can learn to design a typical process plant using simulation software.

Ajay S. Satpute
Process Coach and Managing Director

Converge Engineering Pvt. Ltd.
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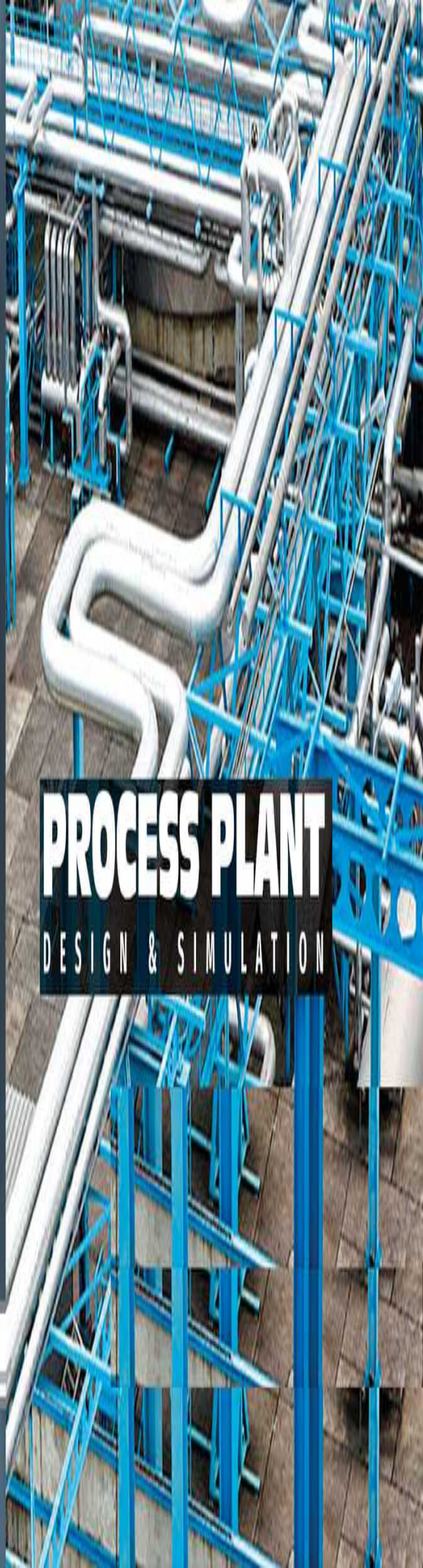

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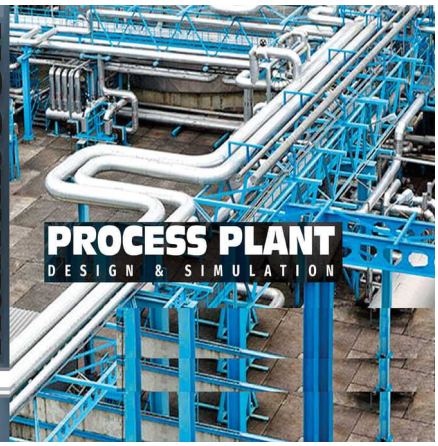

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www.converge.net.in
info@converge.net.in

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ABOUT THE AUTHOR

Mr. Ajay S. Satpute is a reputed name in the Chemical Engineering field with 18+ years' experience in several industries like Oil & Gas, Chemicals, Energy, Utilities, Fertilizers, Iron & Steel, and Pulp & Paper etc. in India, Middle East, and the Far East. He has worked in the process industry in production, technical services and design department.

He has trained several process engineers in the field of process calculations and simulation.

He is an alumnus of the prestigious Indian Institute of Technology (I.I.T.), Madras with Master's Degree in Chemical Engineering.

He has shared several process calculation spreadsheets online. Over 50,000+ engineers worldwide have been benefited by his spreadsheets, articles and his guidance.

Dedicated to all my students, from whom I have learnt a lot.

ACKNOWLEDGEMENT

During the lockdown period of COVID-19 in the first week of April, my wife, Surekha (my boss and director of ***CEPL***), suggested, "Why don't you write a book about plant design and simulation that process engineers can use as a working guide?". Then, the wheels started spinning and I could not sleep that night. Thanks dear, for that push (or kick) that kept me busy for the next few months writing this book.

I'm very proud of my elder daughter, Rhea, for working on spell-check, formatting, drawing schematics & tables, numbering the figures & tables after the draft copy of this book was completed. I'm also thankful to my family members, my mom, sisters, and my younger daughter, Riena for their constant support and encouragement, not just for this book, but during the 20 years of my professional life and even before that.

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Many thanks to my guru, Prof. (Dr.) Dhruvprasad Mishra Sir for reviewing this book and providing valuable suggestions.

I'm grateful to those wonderful seniors, teachers who taught me everything they knew about process engineering. Without them, I would have never known how beautiful process engineering is. Many names come to my mind, Mr. Bhikaji P. Bhayye, Mr. SBVJ Rao, Prof. (Dr.) K. Krishnaiah, Mr. Atul Goswami, Mr. Sanjay B. Patil, Mr. Sudhir Kumar Mittal, Mr. S. Niraipandian, Mr. S. Chinnadurai, Mr. Raj K. Sahni, Mrs. Jayita Sen-Ray, Mr. Sivasubramanian N. and many more.

I remember my (late) dad on this day, who, as usual, would have been very proud of me.

Thanks to the Almighty for everything.

Thx.
Ajay S. Satpute
December 2020

PREFACE

Process engineering, and especially, process design, in my opinion, is the most interesting and beautiful subject, there is. This book is an honest attempt to share the beauty of the subject with everyone. It will certainly help become an excellent process engineer. On purpose, it has been tried to keep the theoretical aspects at bay and focus mainly on practical implications of process design. Once the "how to do" part is clear, then readers will be ready for figuring out the "why" part themselves.

This is a must-have book for final year engineering students and for practicing engineers in engineering consultancies.

It must not be read like a storybook, because it is not one. Rather, it should be referred to while using spreadsheets or software and solve the problems with the help of this book.

I do not claim that this book will provide all the aspects of process design, but I can guarantee that if you follow this book thoroughly, then you will gain enough confidence to carry out any process plant design using simulation software.

It has been a great challenge for me to pen down my experience of designing, running and troubleshooting of the process plant on paper. I hope the younger generation gets benefitted by this.

Before entering process design, one must know what process engineering is. The same has been explained in Chapter 1 (Getting *started*).

Chapter 2 (*Design condition evaluation*) provides different terms and their meaning related to design pressure and temperature.

Chapter 3 (*Process deliverables*) discusses major process deliverables like design basis, simulation or calculation report, operating & control philosophy, line list, datasheet, PFD, P&ID, and C&E matrix.

Nowadays there are several process simulation software which are used in process design. Chapter 4 (*Process simulation software*) provides a brief on different software.

Once the reader has a fair idea about process engineering, deliverables, and software, the actual sizing shall be initiated in Chapter 5 (Line *sizing using Aspen HYSYS*) with line sizing in Aspen HYSYS. Pump sizing is discussed in Chapter 6 (*Pump sizing using Aspen HYSYS*). Aspen HYSYS is used for the simulation.

Chapter 7 (*Control valve sizing using Aspen HYSYS*) provides a step-by-step procedure to carry out control valve sizing using Aspen HYSYS.

Orifice sizing using Aspen HYSYS is discussed in chapter 8 (*Orifice* sizing *using aspen HYSYS*).

Chapter 9 (*Separator sizing using Aspen HYSYS*) discusses separator sizing using Aspen HYSYS.

Chapter 10 (*Heat exchanger* sizing *using Aspen EDR*) discusses heat exchanger sizing using Aspen EDR.

Distillation column sizing using Aspen HYSYS is discussed in chapter 11 (*Distillation column sizing using Aspen HYSYS*).

There is an important tool, Optimizer, in Aspen HYSYS. The same is discussed in chapter 12.

PSV sizing using Aspen HYSYS is carried out in Chapter 13 (*PSV sizing using Aspen HYSYS*).

A real project problem is solved using Aspen HYSYS in chapter 14 (*Process plant design using Aspen HYSYS*).

Chapter 15 (*Aspen HYSYS Dynamics – Simple problem*) discusses a typical Dynamic simulation problem.

In the next chapter (*Aspen HYSYS Dynamics – Project* Problem), the real project problem solved in chapter 14 is converted into a dynamic model and analyzed.

Typical interview questions are listed in the last chapter.

There are several exercises provided. It would help more to the reader if he/she prepares a presentation slides for each exercise and present in front of a group of process engineers.

TABLE OF CONTENTS

CHAPTER 1: GETTING STARTED

Process engineering
What does a Process Engineer do?
Skillset required for Process Engineers
Various packages used in industry
Concept / Feasibility Study
Front End Engineering Design
Detail Engineering Design
As-built drawings
Process Standards
Exercises
CHAPTER 2: DESIGN CONDITION EVALUATION
Temperature System
Operating Temperature
Maximum Operating Temperature
Upset Temperature
Design Temperature
Minimum metal temperature
Pressure System
Operating pressure
Maximum operating pressure
Design pressure
Vapor and vapor-liquid systems protected by relief valves
Vapor and vapor-liquid systems protected by rupture disks
Liquid-Full systems
Shell & tube heat exchangers

Pipeline systems
Lower Design pressure
Maximum allowable working pressure
Maximum allowable incidental pressure
Hydrostatic Test Pressure
Exercise
CHAPTER 3: PROCESS DELIVERABLES
Process Design Basis
Block Flow Diagram
Process Flow Diagram
Piping & Instrumentation Diagram
Operating, Control & Safeguarding Philosophy
Process Datasheet
HAZID
HAZOP
Exercises
CHAPTER 4: PROCESS SIMULATION SOFTWARE
Aspen HYSYS
Aspen Flare System Analyzer
Aspen Exchanger Design and Rating
FluidFlow (from Flite Software NI Ltd.)
Bentley WaterGEMS
Bentley HAMMER
PIPESIM (from Schlumberger)
FlareSim (from Schlumberger)
OLGA (from Schlumberger)
VMGSim (from Schlumberger)

UniSim (from Honeywell)
PIPENET (from Sunrise Systems Ltd.)
Flow Master (from TechNet Alliance)
EPANET (from EPA)
PRO II (from AVEVA)
MySep (from Kranji Solutions)
Pipeline Studio (from Emerson)
Petro-SIM (from Yokogawa)
Exercise
CHAPTER 5: LINE SIZING USING ASPEN HYSYS
Velocity criteria
Pressure drop criteria
Misc. criteria
Water service
Liquid and Gas service
Flow-induced vibrations
Pulsation and transient vibrations
High-frequency acoustic excitation
Transient/Surge analysis
Measures for mitigating surge pressure in pipeline
Simulation
Exercises
CHAPTER 6: PUMP SIZING USING ASPEN HYSYS
Centrifugal Pump
Positive Displacement Pump
PD Pump behavior and safety
PD Pump Types

Minimum Flow By-Pass
Centrifugal Pumps Guidelines
Positive Displacement Guidelines
Capacity Adjustment for Metering Pumps
Pressure Relief for Positive Displacement Pumps
Pulsation Devices for Positive Displacement Pumps
Net positive suction head available
Net positive suction head required
Differential Pressure or Differential head
Pump Motor Brake Kilowatt
Typical Pump Curve
Simulation
Exercises
CHAPTER 7: CONTROL VALVE SIZING USING ASPEN HYSYS
What is a control valve?
What is a Cv?
Control Valve Sizing Criteria
Valve Sizing Criteria
Common flow characteristic curves
Control Valve selection process
Self-acting regulators
Types of self-acting regulators
Simulation
Exercises
CHAPTER 8: ORIFICE SIZING USING ASPEN HYSYS
Simulation
Exercises

CHAPTER 9: SEPARATOR SIZING USING ASPEN HYSYS
Types of Separators
Separator Basic Design Criteria
Mist Extraction Equipment
Souders-Brown Equation
Recommended values of k
Simulation
Exercises
CHAPTER 10: HEAT EXCHANGER SIZING USING ASPEN EDR
Simulation
Exercises
CHAPTER 11: DISTILLATION COLUMN SIZING USING ASPEN HYSYS
Exercises
CHAPTER 12: OPTIMIZER TOOL IN ASPEN HYSYS
Exercises
CHAPTER 13: PSV SIZING USING ASPEN HYSYS
Terminology
When to provide a PRV?
Types of PSV
Chatter
Chatter solution
Rupture disc
quiz
Review of API 520 Part 1
Review of API 521
Review of API 526

Simulation
Exercises
CHAPTER 14: PROCESS PLANT DESIGN USING ASPEN HYSYS
Maximum design pressure
Minimum design pressure
Maximum design temperature
Minimum design temperature
Line sizing criteria
Separator sizing
Atmospheric tank design
Heat exchanger design
Operational drain, vent and flushing requirements
Plant design problem
Simulation
Exercises
CHAPTER 15: ASPEN HYSYS DYNAMICS (SIMPLE PROBLEM)
CHAPTER 16: ASPEN HYSYS DYNAMICS (PROJECT PROBLEM)
CHAPTER 17: TYPICAL INTERVIEW QUESTIONS

CHAPTER 1: GETTING STARTED

Process engineering

Process engineering is the utilization of the laws of physics and chemistry to convert less useful raw materials to more useful products. This involves process plant design, operation, optimization, and troubleshooting. Process engineering is required in almost all industries like agriculture, automotive, biochemical, chemical, material development, mining, nuclear, oil & gas, petrochemical, pharmaceutical, etc.

What does a Process Engineer do?

- Design new equipment/plant as per internationally accepted engineering practices (Greenfield).
- Check the adequacy of the existing system for changed operating conditions (e.g. pressure, temperature, flow, etc.) as per internationally accepted engineering practices (Brownfield).
- Verify new processes.
- Work on optimization projects and troubleshoot plant issues.
- Use process simulation software like Aspen HYSYS or *FluidFlow* to model the process plant.
- Operate the process plant and equipment as per design.
- Size equipment and instrumentation to be implemented and issue relevant Process Datasheets.
- Elaborate Process Flow Diagram and Piping & Instrumentation Diagrams.
- Perform hydraulic and thermal calculations on piping systems, pipelines, flow lines in steady-state and transient scenarios.
- Define the utility requirements and design the associated systems.

- Define the quantity and quality of effluents to be disposed of to the environment.
- Participate in the preparation of the Operating Manual.
- Assess safety and environmental issues.
- Identify process hazards and participate in process safety reviews.

Skillset required for Process Engineers

- Interest in physics, chemistry and mathematics
- Analytical abilities
- Software skills
- Commercial skills
- Multi-tasking skills
- Good communication skills
- Problem-solving skills
- Tenacity

Various packages used in industry

There are mainly 3 packages/studies that are carried out in industry, namely;

- Concept / Feasibility Study
- FEED
- Detail Engineering

Concept / Feasibility Study

This is the beginning stage of any project. Before investing money in any new project, the client needs to ensure that the following concerns are addressed before the investment is justified and the wiser decision can be taken by the client.

- The process must be technically feasible, efficient, and safe.
- Select between batch or continuous plant.
- Budgetary cost of the project.
- Selling plan for the product.

Typical activities during the Concept / Feasibility study

- Evaluation of different process routes w.r.to financial, environmental, and safety points of view.
- Best option recommendation.

- Develop preliminary process flow diagram and heat & mass balance.
- Equipment list.
- Cost estimation (deviation of +50% / -30%).
- Identify weak areas in process design such as experimental design, insufficient laboratory or pilot plant data, scale-up from pilot plant design, and benchmarking against existing designs.

Front End Engineering Design

FRONT END ENGINEERING DESIGN (FEED) is followed by Conceptual design or Feasibility study. FEED design focuses on both, technical and approximate project cost. FEED package forms the basis for the detailed engineering stage.

Typical activities during FEED

- Process Design Basis
- Heat & Material Balance
- Process Flow Diagrams
- Piping & Instrumentation Diagrams
- Line list
- Equipment List
- Hazardous Area Classification Schedule
- Process Datasheets
- Process Description
- Operating, Control and Safeguarding Philosophy
- Process simulation/calculation report
- HAZID / HAZOP study

Detail Engineering Design

All the deliverables prepared during the FEED stage will be reviewed and upgraded during this phase. Many additional documents and drawings specific to this phase are prepared.

Typical activities during Detail Engineering (in addition to FEED)

- Operation & control philosophy
- HAZOP study and Close-out Report
- Pre-commissioning and Commissioning Manual
- Effluents and Emissions Summary
- Single Line Diagrams
- Instrument Loop Diagrams
- Material Requisitions for Static Equipment
- Material Requisitions for Rotary Equipment
- Material Requisitions for Electrical Items
- Material Requisitions for Instrument Items
- Material Requisitions for Civil / Structural
- Piping Key Plan, Piping GA Drawings, Piping Isometric Drawings
- Piping Bulk Material Take-Off (MTO)
- Instrument Bulk MTO
- Electrical Bulk MTO
- Civil Bulk MTO
- Review of Vendor Documents / Drawings (All Disciplines)
- Technical Bid Analysis
- Fire Fighting MTO

As-built drawings

The constructed plant and designed plant do not match exactly all the time. This happens due to many reasons given as below;

- Design stage errors
- Changes in design basis during the construction phase.
- Instructions from the client that led to minimization of equipment/instruments etc.
- On the site changes recommended by commissioning experts during the construction

phase.

EPC contractor provides red mark-up on the P&IDs, Isometric, GADs, etc. post-construction. The red mark-up shall highlight the mismatch between the final design and constructed plant. Engineering consultant updates drawings in CAD and submits the package as “As-Built Drawings”.

Process Standards

Standards are documents that describe the important features of a product, service, or system. Standards are prepared by a professional group or committee which are believed to be good and proper engineering practice, and which contain mandatory requirements and dimensions.

Following major process standards are listed below;

- API 12J : Specification for Oil and Gas Separators
- API 14C : RP for Analysis, Design, Installation, and Testing of Safety Systems for

Offshore Production Platforms

- API RP 14E : RP for Design and Installation of Offshore Production Platform

Piping Systems

- API 520 – I & II : Sizing, Selection, and Installation of Pressure-relieving Devices
- API 521 : Pressure-relieving and Depressuring systems
- API 526 : Flanged Steel Safety Relief Valves for Use in Petroleum Refineries
- ANSI/API 610 : Centrifugal Pumps for General Refinery Service
- API 537 : Flare details for general refinery and petrochemical service
- API 650 : Welded steel tanks for oil storage
- API 674 : Positive Displacement Pumps – Reciprocating
- API 675 : Positive Displacement Pumps - Controlled Volume
- API 681 : Liquid Ring Vacuum Pumps and Compressors for Petroleum, Chemical,

and Gas Industry Services

- API 2000 : Venting Atmospheric and Low-Pressure Storage Tank
- API 2521 : Use of Pressure-Vacuum Vent Valves for Atmospheric Pressure Tanks to

Reduce Evaporation Loss

- ASME B 31.3 : Process Piping Guide
- ASME sec. VIIIB Div. 1: Boiler & pressure vessel code (Design pressure > 3.5 barg)
- ANSI/ISA 75.01.01 : Flow equations for sizing control valves

- EI Guidelines : Energy Institute Guidelines for the Avoidance of Vibration Induced Fatigue

Failure in Process Pipework

- ISGOTT : International Safety Guide for Oil Tankers & Terminals
- ISO 5167-2 : Orifice plates
- OCIMF : Oil Companies International Marine Forum

Exercises

a. Ms. Bindu works in a production company in shift. What would be her job profile?
b. Mr. Satish wants to shift from operating company to engineering consultancy firm. What knowledge he must acquire to work successfully in the new company?
c. Ms. Brinda is a successful process engineer. What skill-set she must have to be successful?
d. Ms. Reena is a billionaire in India. She wants to set up the absolute alcohol production facility in western India. Prepare a concept report with 3 options.
e. Ms. Archana is carrying out FEED for NGL plant in Vizag. What deliverables may be there in the FEED package that she would submit to the end client?
f. Mr. Sanket is a project manager working for an EPC contractor who has been awarded a detail engineering project. What engineering documents do you think he might receive from the engineering consultant?
g. Mr. Gaurav is bidding for a detailed engineering project. Is it wise not to consider any manhours

for As-Built activity?

h. What are the different engineering standards used by a process engineer?

CHAPTER 2: DESIGN CONDITION EVALUATION

The most severe combination of coincident pressures and temperatures normally determines design conditions. In general, piping up to and including DN 600 shall be designed for full vacuum at ambient temperature.

Temperature System

Operating Temperature

The operating temperature is the temperature at which the equipment is operated.

Where equipment operation may be switched from a high-pressure moderate temperature operation to a lower pressure but higher temperature one, then the option of designing for those separate operating modes may be considered (e.g. dual pressure/temperature ratings for the equipment).

Maximum Operating Temperature

The Maximum Operating Temperature (MOT) will be established for operational flexibility and required control system variations. It will be greater than or equal to the operating temperature. The MOT can be used as the basis for materials selection for long term corrosion and/or material degradation.

Upset Temperature

The upset temperature is the maximum temperature that can be reached under upset conditions assuming no barriers are in place. Typical upset scenarios include excessive process heat input (e.g. due to control failure) and loss of cooling medium to the system (e.g. due to power failure). The potential for upset temperatures exceeding the MOT due to exotherms, decompositions, or runaway reactions can be mitigated by one of the following:

- increased design temperature
- instrument safeguards

Fire exposure is an emergency condition that can result in wall temperatures higher than the equipment MOT. As all the process equipment exposed to fire will be subject to inspection, repair, or replacement, fire case temperatures will not be used for establishing design temperature. Upset conditions will be assumed to prevail only for the time required to correct the situation. These conditions will be considered both in materials selection and mechanical design, recognizing that these conditions prevail for a limited period and not indefinitely.

Design Temperature

The design temperature is used for the mechanical design (e.g. determination of minimum wall thickness) of equipment and piping. It is often referred to as the upper design temperature. It is the highest temperature to which equipment is designed at the design pressure. It will be equal to or greater than the MOT. The design temperature is typically at least 10 °C above the MOT.

Design temperature can be rounded up to the nearest 5 °C increment.

For unfired heat exchangers, the design temperature of the cold side can be set equal to the hot side to simplify safeguarding. For tubes in fired equipment, the design temperature can be either the temperature determined by the rules specified above or the calculated maximum tube skin temperature, whichever is higher.

For systems where steam out is intended, the steam out pressure/temperature combination should be specified as a separate design case. Typical steam-out conditions are a temperature of 150 °C under a full vacuum.

For systems where a design temperature has been specified at a value less than 120 °C, the Contractor may elect to designate the design temperature as 120 °C since often there is no increase in flange rating requirement.

Minimum metal temperature

The Minimum Metal Temperature (MMT) is the lowest metal temperature that equipment could reach and will be defined for all equipment, pipelines and piping systems. The MMT will be specified based upon the minimum of the following:

- The lowest normal operating temperature
- The lowest temperature during start-up or shutdown
- The lowest ambient temperature
- The temperature during cool down events caused by high rate depressurization

Pressure System

A pressure system is a single equipment or a group of equipment along with associated piping that within its boundaries remains open under all conditions, and with no possibility of blockage due to freezing, , fouling, sublimation, debris etc. It will not be higher than the pressure rating of any system component. The effects of static head will be considered.

When a system approach is taken, the hydraulics shall be evaluated to ensure that no equipment exceeds its design pressure before a pressure relief device activates, and during relieving conditions, no equipment exceeds the allowable accumulation permitted by either the applicable code or local requirements.

Operating pressure

The Operating Pressure (OP) is the gauge pressure which prevails inside equipment and piping during any intended operation.

Maximum operating pressure

The Maximum Operating Pressure (MOP) is typically at least 105% of the OP to provide adequate flexibility for the control of the intended operations. The MOP may not be less than 1.0 bar above the OP. For pipeline

application Maximum Allowable Operating Pressure (MAOP) or MOP is the maximum internal pressure at which a pipeline system or parts thereof can be operated. MAOP will not exceed 80% of the hydrotest pressure.

Design pressure

The Design Pressure (DP) is the highest gauge pressure that, according to mechanical design codes, an equipment can be continually subjected to at or below the Upper Design Temperature, and above the Lower Design Temperature. It may be rounded up to the next 0.5 barg increment. This pressure provides the basis for relief valve set pressure.

Vapor and vapor-liquid systems protected by relief valves

Equipment that is protected by a relief valve discharging into a flare system or combined vent system shall have a minimum design pressure of at least 3.5 barg due to the potential back pressure exerted on the relief device. When using sonic flare tips, the minimum design pressure will be considerably higher than 3.5 barg. Design pressures lower than 3.5 barg may be considered if the relief valve discharges directly to the atmosphere or is routed to a downstream system that has a design pressure of less than 3.5 barg.

If a DP of 110% MOP is substantially more costly than a DP of 105% MOP, then 105% MOP can be applied, subject to the approval of the client.

The DP is greater than MOP by adequate margin.

For pumps and compressors with an isolation valve between it and upstream system, the suction lines from and including the suction valve shall have the same design pressure as on the discharge side.

The DP of compressor suction and intermediate stage systems (e.g. coolers, condensers, knock-out drums) will be sufficiently high to prevent the opening of pressure

safety valves through pressure equalization “settle out” after the compressor has shut down.
Design pressure and temperature should be optimized to eliminate and reduce relief cases where practically possible, to provide an inherently safe and sustainable unit design, and to reduce costs.

Vapor and vapor-liquid systems protected by rupture disks

Rupture disks are highly stressed in service. To avoid premature failure because of corrosion, creep and fatigue, a substantial margin (at least 30% of the normal rupture pressure for tension-loaded disks, and at least 10% of the minimum rupture pressure for reverse-buckling disks) will be allowed between the MOP and the design pressure of the equipment.

Liquid-Full systems

Liquid-Full systems will be designed to withstand pump shut-off pressure, if they can be “blocked in” while pumps feeding them could continue to operate. The shut-off pressure is the pressure at the discharge of a centrifugal pump with the suction pressure at the maximum possible value and the discharge system closed.
For systems that do not normally operate liquid full but could become liquid fill during an upset condition, it may be determined whether the system is to be designed for shut-in pressure or whether some other means of safeguarding is required.
For steam or gas turbine-driven pumps and variable speed electric motor driven pumps, the maximum possible speed during operation of the pump shall be evaluated since this can influence the design pressure considerably.

Shell & tube heat exchangers

The most unfavorable combination of design pressures on the shell and tube side that occur, will be used in the calculations for tube sheets, floating heads and tubes.

The design pressure of the high-pressure side of the exchanger should be lower than the corrected hydrotest pressure of the low-pressure side of the exchanger or a relief device is required to protect against tube rupture.
If design pressure is specified to eliminate the need for a relief device for tube rupture, the (mechanical) design pressure of the low-pressure side of the exchanger will be extended up to and including the block valves.

Pipeline systems
The design pressure at any point will be equal to or greater than the MAOP.

Lower Design pressure
The Lower Design Pressure (LDP) is the external design pressure or the sub-atmospheric pressure at the top of the equipment in its operating position.
LNG rundown, loading, and circulation lines, as well as boil-off gas lines, will be designed for full vacuum to enable vacuum drying during commissioning.
Vessels that are in steam service should be designed for full vacuum at 150 °C.
For vapor with an atmospheric dewpoint of 0 °C or above, it will be checked whether vacuum can develop in the equipment if the heat input to the equipment fails.
If the pressure drops below atmospheric pressure due to utility failure, instrument failure, etc., the minimum possible absolute pressure shall be stated in the documents for mechanical design.
Steaming-out and draining operations shall be evaluated when determining the LDP.

Maximum allowable working pressure
The Maximum Allowable Working Pressure (MAWP) is the maximum gauge pressure permissible at the top of the equipment with the equipment installed in its operating position and at a designated temperature. The MAWP is equal to or greater than the DP.

Maximum allowable incidental pressure

Maximum Allowable Incidental Pressure (MAIP) is a term used to describe infrequent short periods during which the pipeline system may be subjected to pressures greater than MAOP.

Examples of incidental pressure include surge, failure of pressure controlling equipment, and cumulative pressure during activation of overpressure protection devices. The MAIP should not exceed 90% of the hydrotest pressure.

Hydrostatic Test Pressure

The test pressure shall be 1.5 times the design pressure. Pressurizing of the system shall be carried out in increments of 25% total pressure value or as designated by the Client at the time of testing up to the test pressure. There will be a hold for 15 minutes at each increment. This is to ensure that the system is stabilized and to avoid over pressurizing. Once the test pressure has been attained, it is to be maintained for one hour before the detailed inspection takes place. The test medium shall be potable water.

Exercise

a. Mr. Umesh has a presentation next week to explain different terms related to pressure and temperature used in equipment design. You are his subordinate. Prepare a 10-slide presentation for your boss that will impress your boss' boss.

CHAPTER 3: PROCESS DELIVERABLES

Major process deliverables are discussed in this chapter.

Process Design Basis

The process design basis document is the collection of all the input parameters considered for process design. The process parameters include feed composition, flow rate, operating and design conditions, velocity criteria, pressure drop criteria, process scheme, different process scenarios etc. The client must approve a process design basis before the designer starts any calculations or simulation for process design.

Typically, this document is prepared by process engineer, checked by a senior engineer, and approved by process HOD.

Any change in process design basis will have a major impact on several documents viz. PFD, P&ID, Process calculations/simulation report, Equipment List, Heat Exchanger Thermal Design, Equipment Datasheets, Relief Valve Sizing & Datasheets, Electrical Load List etc.

Each document must undergo stringent quality checks. It starts with self-check. The person generating the document will first self-check his/her document. The same document will then be checked by senior discipline engineers. The document may also be checked by engineers from other disciplines. This checking step is called as Inter-discipline Check. The comments/corrections will be discussed amongst the project team and if those are agreed by the originator, then the same will be incorporated. If there is disagreement, then the project manager will make the final call.

A checklist is a list that guides the engineers to perform a check of the quality of the document and ensure that major criteria for the design are covered. A checklist will be specific to a specific type of project. Many companies have their checklist form specific to their projects.

A typical checklist for the process design basis document is given below;

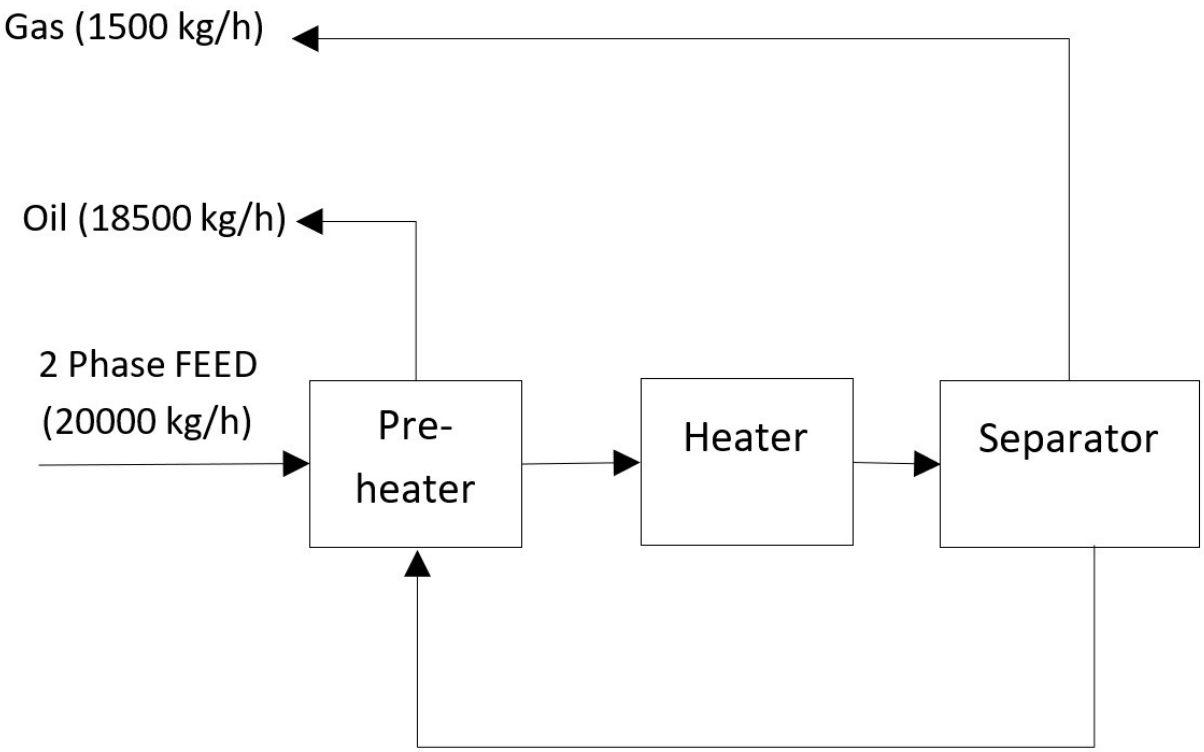
- Correct document titles and revision
- Document number included
- Consultant standard title block and job number
- Consultant Disclaimer description block
- All design information included.
- Operating and design conditions specified
- Requirements for references and are they specified
- Self-Discipline Check comments incorporated
- Inter Discipline Check commented incorporated
- Clients commented incorporated

Block Flow Diagram

A block flow diagram (BFD) is a process drawing used to simplify the basic structure of a system. It is the simplest form of the flow diagrams used in the industry. Blocks in a block flow diagram represent anything from a single piece of equipment to an entire plant.

A typical example is given below;

Figure 3.1: Block diagram



Process Flow Diagram

A process flow diagram (PFD) is a diagram commonly used in engineering to indicate the general flow of the plant process. The PFD displays the relationship between major equipment of a plant facility and does not show minor details such as piping details and designations.

Process flow diagrams of a single unit process will include the following:

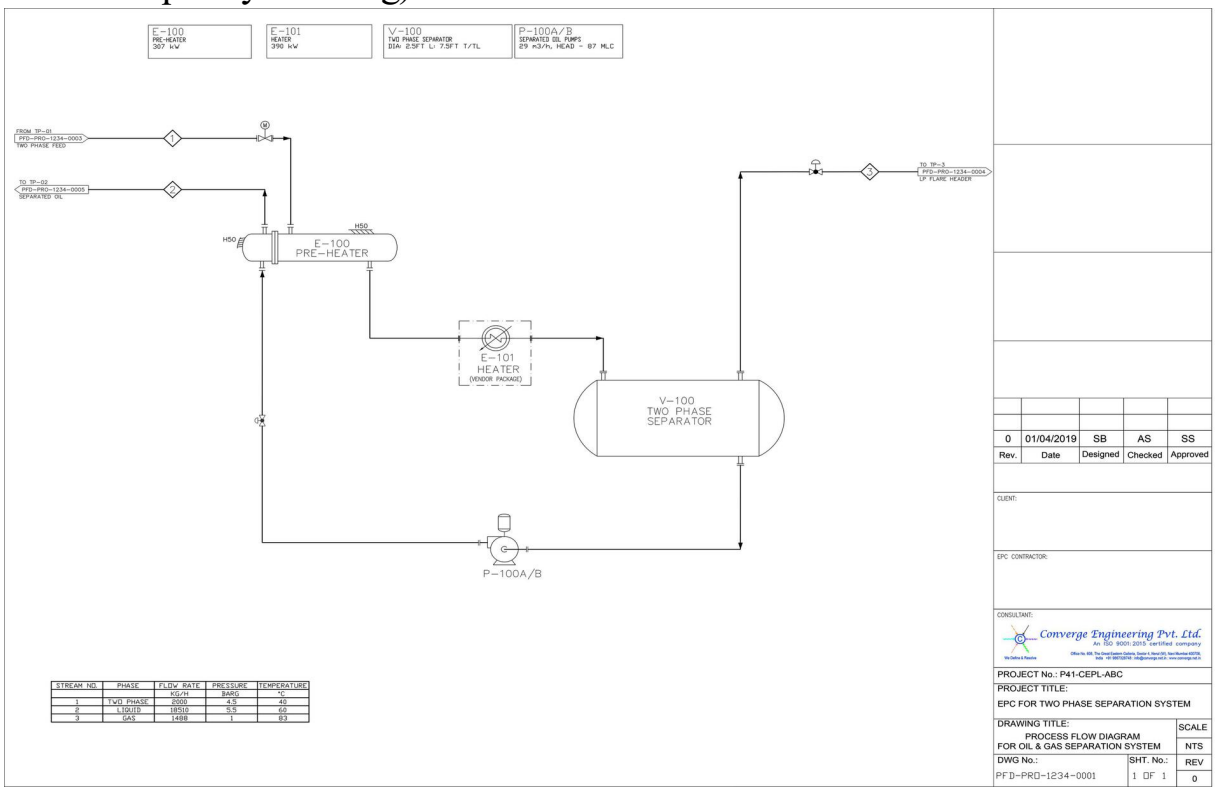
- Major process piping
- Major bypass & recirculation lines
- Major equipment symbols, names, and identification numbers
- Process flow directions
- Major control loops
- Interconnections with other systems
- Stream Identification
- Heat & Mass Balance

PFDs generally do not include the following:

- Piping class or piping line numbers
- Pipe size
- Instrumentation
- Minor bypass lines
- Isolation and shut-off valves
- Maintenance vents and drains
- Relief and Safety Devices
- Pipe fittings and flanges

Typical mass balance and PFD schematic are shown below;

Figure 3.2: PFD (Refer https://cutt.ly/ohSKS5P for better quality drawing)



A typical checklist for process flow diagram is given below;

General

-

Correct drawing titles and revision

- Drawing number included
- Consultant standard title block and job number
- Consultant Disclaimer description block
- Drawing continuation flags match before drawing
- Scope of the project clearly defined by clouding (but use of clouding kept to a minimum)
- Original Master Drawing Number listed
- Specific notes included and numbers matching
- In line with standard drawing legend
- SDC comments incorporated
- IDC commented incorporated
- Clients commented incorporated

Equipment and Packages

- All equipment (operating and standby) included
- Battery limits shown for packages
- Equipment and Packages names and numbering
- Design information data included and as per datasheet
-

Directional arrows on major process and utility lines
- Line sizes shown on major lines

Process Control

- Control valves and selection (pneumatic, motor, etc.)
- Control valve failure positions
- All control loops shown including any logic
- All key Instrumentation shown and tagged
- Control logic as per Operating & Control Philosophy

Safeguarding

- Overpressure protection
- Process Shutdown Valves (PSD)
- Emergency Shutdown Valves (ESD)
- HP/ LP Interfaces
- Non-Return valves
- Emergency blowdown requirements

Miscellaneous Technical

-

Stream numbers on major process streams

- Stream numbers match with Process Model
- Heat and Mass Balance Table
- Future facilities (shown dotted if required)
- Revision triangles and Revision clouds shown on drawing updates

Piping & Instrumentation Diagram

A Piping & Instrumentation Diagram (P&ID) is a pictorial representation of the piping, equipment, instrumentation, and control devices. P&ID preparation is initiated during FEED and is further updated during detail engineering. P&IDs are updated w.r.to vendor data, design review comments, HAZOP recommendations, and as-built changes.

P&ID shows the interconnection of process equipment and the instrumentation used to control the process. In the process industry, a standard set of symbols is used to prepare drawings of processes. The instrument symbols used in these drawings are generally based on the International Society of Automation (ISA) Standard S5.1.

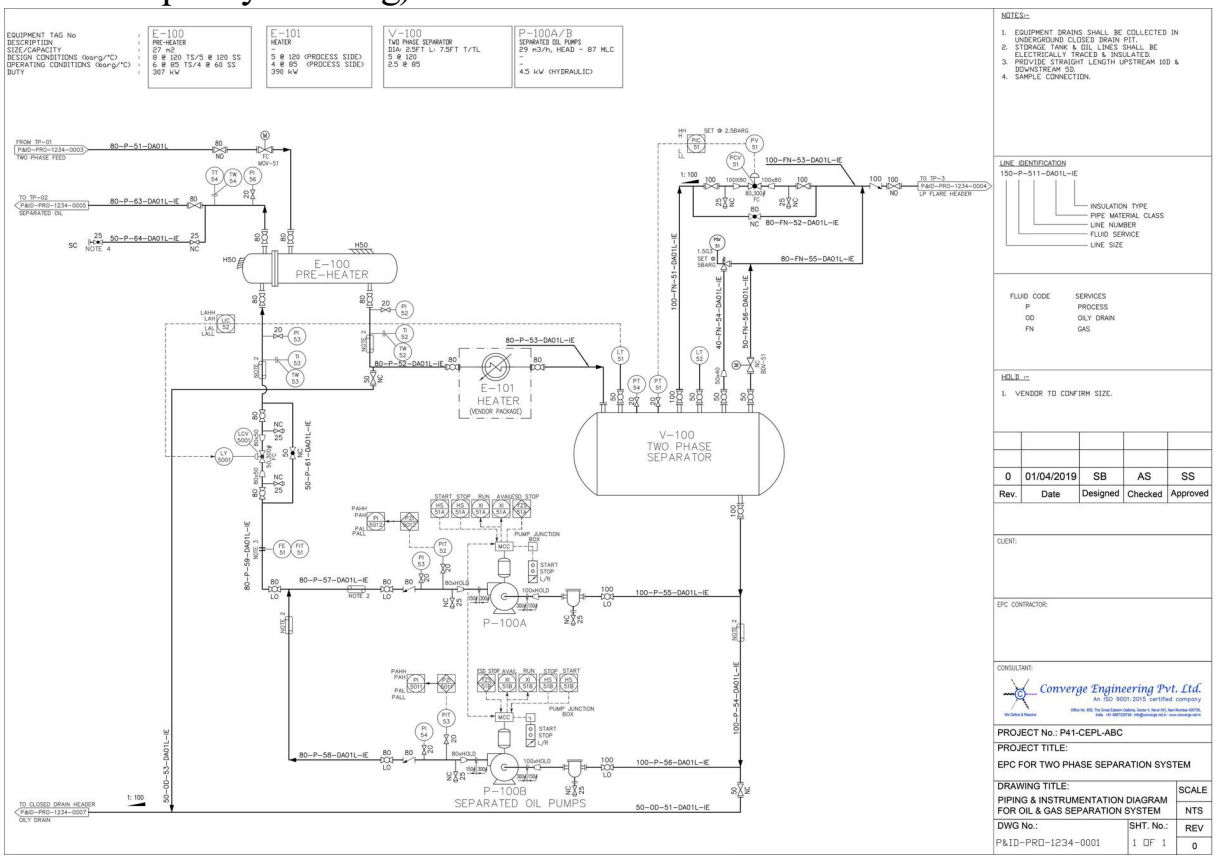
- CAD system is used to develop the P&IDs.
- Most details that are available from other types of documentation (e.g., instrument loop diagrams, vessel data sheets) are not recommended for inclusion on P&IDs.
- P&ID shows equipment with simple outline representation.
- Equipment is not drawn to scale, however equipment relative to one another both in size and general orientation must be maintained.
- P&ID shows nozzles on equipment, including spares, as single lines.
- P&ID indicates nozzle sizes unless the size is implied by piping connections.
- P&ID shows equipment item number and title/service as a minimum.
- P&ID indicates internals for equipment as dashed lines. Details of internals that have no significant bearing on the piping design and layout or equipment operation are omitted.
- P&ID indicates auxiliary system requirements for individual pieces of equipment (e.g. lube oil

systems, seal flush systems, turbine gland leak-off piping, sample systems) on auxiliary P&IDs.

- P&ID indicates jacketing requirements for equipment and jacketing/tracing requirements for piping.
- P&ID indicates the type of insulation (e.g., personnel protection, heat conservation) for equipment as part of the equipment data.
- P&ID indicates insulation thickness where applicable.
- P&ID indicates a piping serial number and piping class.

A typical P&ID is shown below;

Figure 3.3: P&ID (Refer https://cutt.ly/ohSKS5P for better quality drawing)



A typical checklist for Piping & Instrumentation diagram is given below;

General

- Correct drawing titles and revision
- Drawing number included
- Consultant standard title block and job number
- Consultant Disclaimer description block
- Construct and Demolition drawing numbers match
- Drawing continuation flags match before drawing
- Scope of the project clearly defined by clouding (but use of clouding kept to a minimum)
- Original Master Drawing Number listed
- Specific notes included and numbers matching
- In line with Standard drawing legend
- SDC comments incorporated
- IDC commented incorporated
- Clients commented incorporated

Equipment

- All equipment (operating and standby) included
- Equipment names and numbering shown
- Equipment design information data included and as per datasheet
- Equipment and internals/externals as per datasheet
- Nozzles, manways and numbering as per datasheet
- Insulation and Trace Heating
- Draining, Venting and Purging
- Equipment Isolation including valves and spades
- Overpressure protection, relief valves, bursting discs
- Equipment trim numbers

Packages

- Package names and numbering
- Package information data blocks completed
- Piping & Instrument Connections as per datasheet
- Battery limits shown
- References to Suppliers PEF'S included

Piping

- Directional arrows on major process & utility lines
- Line sizes on all lines
- Line numbers on all lines including pipe specs
- Requirement for slopes and avoiding pockets shown
- Tie-in Point numbers including spec breaks
- Tie-in points between Construct and Demolition drawing match
- Insulation and Trace Heating
- Pipe specifications breaks shown
- Piping special items included and numbered
- Overpressure protection, TRV's, bursting discs
- Valve selection in line with Pipe specs
- Sample points and corrosion monitoring

Process Control and Instrumentation

- Correct control valves selection (pneumatic, motor, etc.)
- Control valves sizes, tag numbers & failure positions
- Control valve isolation block valves & drains/vents
- Control valve bypasses (to be avoided where possible)
- All control loops shown including any logic

- All Instrumentation shown and as per PFS including tag numbers
- Alarm and trip setpoints
- Vessel level alarm and trip setpoints
- Tapping sizes on equipment/piping

Safeguarding

- Relief valves and numbering
- Relief valve set pressures and line sizes
- Relief valve sparing, isolation and interlocks
- Emergency shutdown valves and numbering
- Non-Return valves
- Emergency blowdown requirements

Line List:

A line list is a database that has the following information for process and mechanical engineering teams.

One such example is given below.

Figure 3.4: Line list

10-6" -WW-5765-A-040-HC	0-6" -WW-5001-A-040-HC-ST	**Line ID**	
10-6" -WW-5645-A-040-HC	10-6" -WW-5004-A-040-HC-ST	**Start**	
V-6000	10-6" -WW-5008-A-040-HC-ST	**End**	
Waste Water	Waste Water	**Service**	
Liquid	Liquid	**Phase**	
15	15	**Pressure, barg**	**Design**
150	120	**Temperature, °C**	
3	5	**Pressure, barg**	**Operating**
55	55	**Temperature, °C**	
1"	1.5"	**Insulation**	
N/A	Steam	**Heat Tracing**	
PD-5678	PID-5678	**P&ID**	
ISO-6087	ISO-5992	**Isometric Drawings**	
		Notes	

A typical checklist for line list is given below;

- Correct drawing titles and revision
- Drawing number included
- Consultant standard title block and job number
- Consultant Disclaimer description block
- Standard document front sheet included
- Standard Hold List and action party (in brackets)
- All lines/equipment included as per P&ID/PDS
- All lines/equipment numbering matches P&ID/PDS
- All design information included and as per P&ID/PDS
- Operating and design conditions specified
- For line list, “from” and “to” locations match with P&ID
- Requirements for references and are they specified
- Check that references to P&ID/PDS match with P&ID
- SDC comments incorporated
- IDC commented incorporated
- Clients commented incorporated

Operating, Control & Safeguarding Philosophy

Operating, control & safeguarding philosophy (OCSP) is a document that covers the following aspects;

- The operating philosophy of the plant/unit is described in a detailed manner. The mode and sequence of operations are mentioned. Equipment description, tag numbers, normal operating parameters of flow, pressure, and temperature, and stand-by equipment availability may be mentioned in this section.
- Control Philosophy describes all controls required for the safe, reliable, and uninterrupted operation of the plant/unit. This could include flow, pressure, and temperature and level control of the plant/unit for the smooth operation of the plant/unit. Controls required for start-up, planned shutdown and to change plant/unit capacity should be mentioned. High and Low alarms for process operating parameters are also described in this section.
- Safeguarding Philosophy describes the safeguarding functions that may be available for the prevention of the uncontrolled process, risk reduction or mitigation in case of external hazards, manually initiation of an emergency shutdown of the plant/unit by operator intervention.

Typically, this document is prepared by process engineer, checked by a senior engineer, and approved by process HOD.

Any change in Operating & control philosophy will have a major impact on the Cause and Effect diagram.

The checklist will be similar to the one given in the earlier section.

Process Datasheet

The process data sheet PDS relates to a single item of equipment and contains the essential process data for initiating the detailed design of an item. It includes the overall size, number, approximate geometry and identification of the connections, material of construction, and the full range of operating conditions. The process data sheet generally includes a simple diagram. Also, the process design includes Instrument process datasheets which provide a similar level of detail.

Refer to the example below:

Figure 3.5: Process datasheet

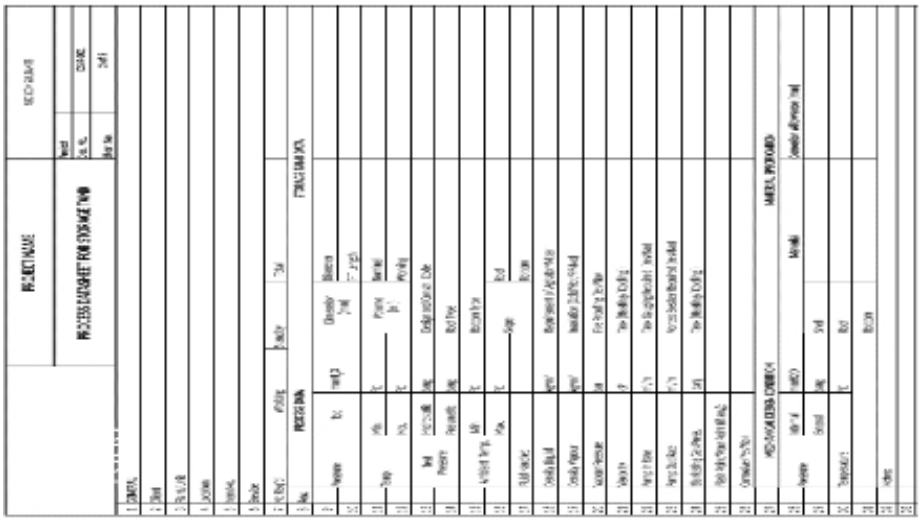
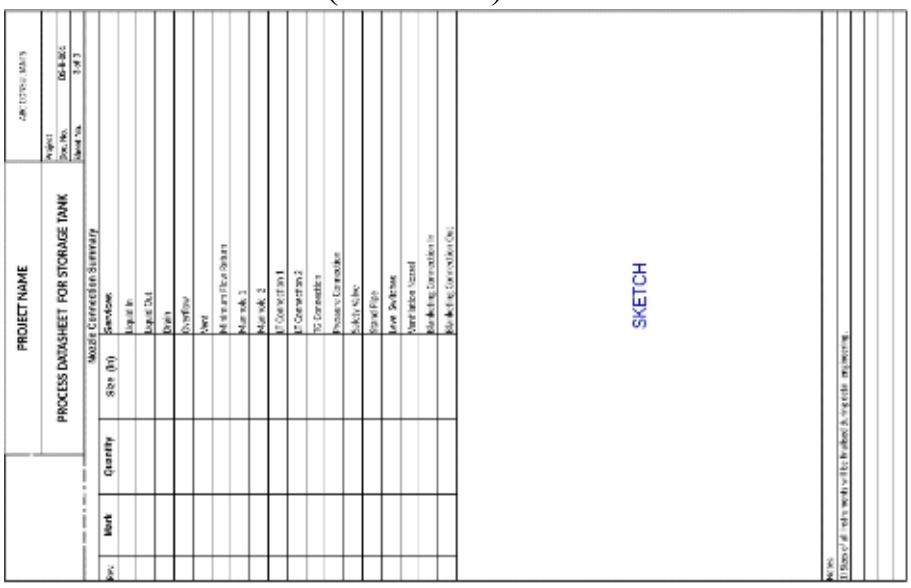


Figure 3.6: Process datasheet (continued)



HAZID

HAZID (Hazard Identification) is a well-documented qualitative technique for the early identification of potential hazards and threats affecting people, the environment, assets or reputation. The major benefit of a HAZID study is to provide essential input to project development decisions. It is a means of identifying and describing HSE hazards and threats at the earliest practicable stage of development or venture. HAZID is used both as part of a Quantitative Risk Assessment and as a standalone analysis for i.e. installation, modification, replacement, upgrading, reduction, isolation and lifting etc.

HAZID Study Objectives

- Identify the potential hazards.
- Reduce the probability and consequences of an incident at a site that would harm the personnel, plant, and environment.
- Identify opportunities for inherent safety.
- Identify Fire, explosion, toxic release scenarios, and measure to prevent it.
- Hazards involved in operating each equipment can be enlisted at the beginning.
- Early assessment of health, safety, and environmental hazards.
- Provide essential input to the project development decisions.

HAZOP

HAZOP (Hazard and Operability) study is done to review a process or operation systematically to determine whether deviations from the design can lead to undesirable consequences. HAZOP can be used for continuous or batch processes and can be adopted to evaluate written procedures. The HAZOP team lists potential causes and consequences of the deviation as well as existing safeguards protecting against the deviation. When the team determines that inadequate safeguards exist for a credible deviation, it usually recommends the action to be taken to reduce the risk. The minimum documents that are needed to carry out HAZOP are Piping & Instrumentation Drawing, Cause & Effect Matrix, and Operating & Control Philosophy. HAZOP session is attended by the client (process, production, maintenance dept.), EPC contractor (Project manager, HSE engineer), Design Consultant (Process, Instrumentation), Vendor (if any) and chaired by a third-party safety consultant (called as HAZOP Chairman).

HAZOP Study Objectives

- To assess the hazard potential of malfunction of equipment and its impact on the facility.
- To brainstorm how a process may deviate from the design intent.
- Identify potential problems w.r.to design. The solution to those problems may or may not be found in the HAZOP session.
- Evaluate and question the purpose of every control or safety items given in P&IDs.
- To recommend adequate changes in the design based on the HAZOP session participants' agreement.
- To improve the safety of existing facilities

- To evaluate and minimize risks w.r.to personnel, environment, material, and reputation.

Exercises

a. Mr. Yashpal is a fresher in a reputed consultancy in Navi Mumbai. As his first assignment, his boss has asked him to prepare a process design basis for a cooling water supply system project. Can you advise Mr. Yashpal on how to prepare such a document?
b. Ms. Sayali needs to prepare a block diagram and PFD of the BTX separation system. How would she include in both the drawings?
c. Mr. Aditya is having trouble with the BTX system P&ID. Can you help him draw such a system?
d. Mr. Nimesh is assigned to prepare a line list for Mr. Aditya's project. Can you help him with this activity?
e. Mr. Saurabh is also a part of your team who is supposed to complete OCSP of BTX plant. He has been reassigned to some other urgent project. You are his replacement. Please submit his deliverable.
f. Mr. Pradip is preparing a cooling water pump datasheet. Can search on the internet and help him with the template.
g. Mr. Prasad is the HAZID chairman. What exactly are his job responsibilities?
h. Mr. Naresh is planning to conduct a HAZOP session for the BTX plant designed by Mr. Aditya. How exactly he will execute this activity?

CHAPTER 4: PROCESS SIMULATION SOFTWARE

Process simulation software is used for the design, development, analysis, and optimization of technical processes such as chemical plants, chemical processes, environmental systems, power stations, complex manufacturing operations, biological processes, and similar technical functions. There are several commercial simulation software available. A few of those are discussed below.

Aspen HYSYS

General process simulation and modeling software very commonly used in the upstream oil and gas industry is HYSYS. HYSYS is extensively used for developing PFD and mass and heat balance.

Aspen Flare System Analyzer

It is extensively used for modeling of flare or cold vent networks commonly required in upstream oil and gas. It allows determining the type of relief device and sizing of connected piping including individual tailpipes, flare sub-header, and flare the main header.

Aspen Exchanger Design and Rating

Aspen Exchanger Design And Rating (EDR) enables you to easily design and simulate the heat Exchangers. Deliver the optimum heat exchanger size and rating when both cost and performance must be balanced with rigorous up-to-date integrated modeling capabilities.

FluidFlow (from Flite Software NI Ltd.)
FluidFlow is the most comprehensive and easy to use pipe flow software. It has an intelligent and easy-to-use graphical interface. FluidFlow has following modules:

- FluidFlow Liquid
- FluidFlow Gas
- FluidFlow Two-Phase
- FluidFlow Non-Newtonian & Slurry

Bentley WaterGEMS
It is used for the analysis and design of water networks. Create and manage customized reports that automatically combine graphs, data tables, color-coded and annotated plan views, and more into a single report. Fire flow, hydraulics, operations, criticality, and water quality can be modeled.

Bentley HAMMER
HAMMER determines appropriate surge control strategies and reduce transients. It uses the Method of Characteristics, the benchmark standard for hydraulic transient analyses. One can run both transient and steady-state analyses in HAMMER.

PIPESIM (from Schlumberger)
A simulation and modeling software used for steady-state modeling of "pipelines" both for single-phase, two-phase and three-phase flow.

FlareSim (from Schlumberger)
Flaresim provides full 3D flame-shape analysis with complete flexibility in the location and orientation of multiple stacks, and rapid evaluation of flare systems under different wind speeds and directions. Shield options can be included. Working from the opposite perspective, Flaresim allows you to design stack or boom length to

meet specific radiation, noise, or surface-temperature limits at defined receptor points.

OLGA (from Schlumberger)

The OLGA dynamic multiphase flow simulator models time-dependent behaviors, or transient flow, to maximize production potential. Transient modeling is an essential component for feasibility studies and field development design. Dynamic simulation is essential in deepwater and is used extensively in both offshore and onshore developments to investigate transient behavior in pipelines and wellbores.

Transient simulation with the OLGA simulator provides an added dimension to steady-state analyses by predicting system dynamics such as time-varying changes in flow rates, fluid compositions, temperature, solids deposition, and operational changes.

From wellbore dynamics for any well completion to pipeline systems with all types of process equipment, the OLGA simulator provides an accurate prediction of key operational conditions involving transient flow.

VMGSim (from Schlumberger)

VMGSim is a general process simulator and possesses a complete suite of process unit operations which can be configured to model virtually any process of interest in the Gas processing, Refining/Petrochemical or Chemical industries.

It is an advanced and fully interactive steady-state process simulator that offers a modern and integrated flowsheet environment, letting you study your plant in sections or as a whole. It tracks all of your data inputs, making it possible to immediately see what was added to the system and the result.

UniSim (from Honeywell)

UniSim Design Suite is an intuitive process modeling software that helps to create steady-state and dynamic

models for plant design, performance monitoring, troubleshooting, business planning, and asset management. Its major use:

- Process flowsheet development
- Utilizing case scenarios tool to optimize designs against business criteria
- Equipment rating across a broad range of operating conditions
- Evaluating the effect of feed changes, upsets, and alternate operations on process safety, reliability, and profitability
- Accurately size and select the appropriate material for blowdown systems
- Monitoring equipment performance against operating objectives.

PIPENET (from Sunrise Systems Ltd.)
PIPENET is the leading software for a rapid fluid flow analysis of pipe and duct networks. Three modules (Standard Module, Spray/Sprinkler Module, Transient Module) ensure that, no matter how complex your network, PIPENET software will perform flow simulation fast and accurately.

Flow Master (from TechNet Alliance)
It is a 1D Computational Fluid Dynamics (CFD) software that allows thermo-fluid system engineers to model and analyze the fluid mechanics and pipe flow in complex systems.

EPANET (from EPA)
It is a public domain, water distribution system modeling software package developed by the United States Environmental Protection Agency's (EPA) Water Supply and Water Resources Division. It performs an extended-period simulation of hydraulic and water quality behavior within pressurized pipe networks.

PRO II (from AVEVA)

It is a steady-state process simulator for process design and operational analysis for process engineers in the chemical, petroleum, natural gas, solids processing, and polymer industries. It includes a chemical component library, thermodynamic property prediction methods, and unit operations such as distillation columns, heat exchangers, compressors, and reactors as found in the chemical processing industries. It can perform steady-state mass and energy balance calculations for modeling continuous processes.

MySep (from Kranji Solutions)
MySep enables the user to effectively determine vessel sizes, select internals, and assess the associated overall performance of separation vessels. It can also be used to evaluate existing designs and determine theoretical performance based on actual or simulated operating data.
MySep can be used independently and in conjunction with process modeling software as Aspen HYSYS, UniSim Design, Petro-SIM, VMGSim, and PRO II.

Pipeline Studio (from Emerson)
Pipeline Studio is the industry-leading pipeline design and engineering solution that combines graphical configuration and reporting tools with industry-proven simulation engines. It provides fast, accurate, robust, and reliable answers to a wide range of steady-state and transient analysis challenges.
Pipeline Studio delivers rapid and accurate offline design, planning, and hydraulic analysis for natural gas and liquid pipelines through advanced state-of-the-art simulation techniques. The combination of both steady-state and transient hydraulic simulation within a feature-rich Windows graphical interface enables a better understanding of even the most demanding problems by providing an appropriate analysis of the process.
Use Pipeline Studio for:

- Flow assurance
- Designing, routing, sizing of pipeline networks
- Upset, leak and survival time analysis
- Strategic, operational and capacity planning
- Rapid assessment of unscheduled changes in operation
- Operational prognosis
- Compressor requirements
- Fuel consumption calculations
- Line pack management

- Assessment of storage requirements
- Surge Analysis

Petro-SIM (from Yokogawa)

Petro-SIM is a process simulation software and optimization platform for driving excellence in facility performance and organizational productivity. At the core of Petro-SIM's technology are rigorous process simulation models that generate dependable results in an intelligent, user-friendly environment. Petro-SIM is the go-to simulation platform for upstream, midstream, and downstream refinery and petrochemical processes. For upstream oil and gas, Petro-SIM is the only purpose-built process simulation software that can ensure facility processing capabilities meet long term reservoir needs.

In oil refining and petrochemicals, our Petro-SIM process simulation software works in tandem with our world-leading SIM Reactor Suite to create the only proven simulator capable of scaling from reactor units to entire facilities.

Exercise

a. Ms. Sushama needs to present in the next process department knowledge sharing meeting. She has been given a topic as different process simulation software. Please help her with this activity.

CHAPTER 5: LINE SIZING USING ASPEN HYSYS

Line sizing is perhaps the most widely used process calculation activity and the most important as well. Line sizing means to find out the adequate line or pipe size for the given flow rate. There are mainly 2 line sizing criteria; velocity criteria and pressure drop criteria.

The sizing criteria for liquid piping systems will depend on the application. The function and application of the piping system will determine the sizing criterion to be selected. Where pressure drop is not a determining parameter, the size should be determined by the velocity constraints.

The velocities shall be kept low enough to prevent problems with erosion, water hammer, pressure surges, noise, and vibration, and reaction forces. In some cases, a minimum velocity is required. When determining the velocity of the medium in the lines, account should be taken of the possible generation of static electricity.

Velocity criteria

- Pump suction line (Typical velocity ~ 1 m/s)
- Pump discharge line (Typical velocity ~ 3 m/s)
- Gas line (< 18 m/s)
- Gravity line (~ 1 m/s)
- Two-phase line (V < Ve)

Pressure drop criteria

- Sub-cooled liquids (0.25 bar / 100 m)
- Boiling Liquids (0.05 bar / 100 m)

Misc. criteria

- PSV inlet line ($\Delta P < 3\%$ of set point pressure)
- PSV outlet or tailpipe (Mach no. < 0.7)
- Flare header (Mach no. < 0.5)
- Flare lines ($\rho V^2 < 200{,}000$ kg/ms^2)
- Depressurization line (Mach no. < 0.7)

- Vent line for atmospheric tank (MABP < 0.07 bar)

When sizing piping, the following constraints shall be addressed:

- Required Capacity / Available Driving Pressure
- Noise / Vibration
- Pressure Surges
- Material Degradation – Erosion, Corrosion, Cavitation
- Solids Accumulation

Pipe Roughness

For all calculations of pressure drop, the following pipe roughness values should be used:

- Carbon steel (CS) corroded: 0.46 mm (0.018 inch) (Note-1)
- Carbon steel (CS) non-corroded (for relief system piping): 0.15 mm (0.006 inch)
- Carbon steel (CS) non-corroded (other systems): 0.046 mm (0.0018 inch)
- Stainless steel (SS): 0.046 mm (0.0018 inch)
- Titanium and Cu-Ni: 0.046 mm (0.0018 inch)
- Glass fiber reinforced polyester (GRP): Vendor to provide
- Polyethylene, PVC: Vendor to provide

Note 1 – The value of 0.46 mm (0.018 inch) shall be used when hydraulic calculations are performed for existing pipe installations including relief piping.

Water service

Hazen-Williams equation is used for pressure drop calculation.

The Hazen–Williams equation is an empirical relationship that relates the flow of water in a pipe with the physical properties of the pipe and the pressure drop caused by friction. It is used in the design of water pipe systems such as fire sprinkler systems, water supply networks, and irrigation systems. It is named after Allen Hazen and Gardner Stewart Williams.

The Hazen–Williams equation has the advantage that the coefficient "C" is not a function of the Reynolds number, but it has the disadvantage that it is only valid for water. Also, it does not account for the temperature or viscosity of the water.

$$h_f = 10.67\, L \frac{Q^{1.85}}{C^{1.85}\, d^{4.87}}$$

Where;

- h_f = head loss in meters (water) over the length of pipe
- L = length of pipe in meters
- Q = volumetric flow rate, m^3 /s (cubic meters per second)
- C = pipe roughness coefficient
- d = inside pipe diameter, m (meters)

Liquid and Gas service

The pressure loss in pipe flow is calculated using Darcy-Weisbach equation (equation 1).

$$\Delta P = \frac{f_D L \rho V^2}{2d}$$ Equation 1:

This equation is valid for both laminar and turbulent flow regimes of any fluid with fully developed and incompressible flow. It can be used for gas services as long as there is only a negligible change in gas density across the pipe length. The formula for calculating the Darcy friction factor is different for laminar (equation 2) and turbulent flow (equation 3 - Haaland equation).

Laminar flow meaning Reynolds number less than 2000.

Equation 2:

$$f_D = \frac{64}{N_{Re}}$$

Equation 3:

$$\frac{1}{\sqrt{f_D}} = -\frac{1.8}{n}\log\left[\left(\frac{6.9}{N_{Re}}\right)^n + \left(\frac{\epsilon}{3.75\,d}\right)^{1.11n}\right]$$

The Haaland equation is used to solve directly for the Darcy–Weisbach friction factor f_D for a full-flowing circular pipe. It is an approximation of the implicit Colebrook–White equation, but the discrepancy from experimental data is well within the accuracy of the data.

Where;

- ΔP : Pressure drop, Pa
- f_D : Darcy friction factor
- L : Length of pipe, m
- V : Velocity, m/s
- N_{Re} : Reynolds number
- ϵ / d : Relative roughness
- d : Inside pipe diameter, m
- ρ : Liquid density, kg/m^3
- n : Constant (n=1 for liquid services and n=3 for gas services)

Flow-induced vibrations

Flow-induced vibrations are more predominant in turbulent flow regime piping systems. The turbulent flow regime generates potentially high levels of kinetic energy local to the turbulent source. This energy is distributed across a wide frequency range, most of the excitation is concentrated at low frequency. This type of vibration leads to displacement of the piping system and in some cases also leads to damage to pipe supports.

Susceptibility to failure of piping systems due to flow-induced vibrations:

Table 5.1:

Aspect	Fluid phase	Classification for susceptibility to failure		
		Negligible	Medium	High
Kinetic energy of piping system	Liquid and multi-phase	$\rho v^2 < 10000$ kg/ms^2 ($\rho v^2 < 6700$ lb/fts^2)	10000$< \rho v^2 <$ 20000 kg/ms^2 (6700$< \rho v^2 <$ 13400 lb/fts^2)	$\rho v^2 \geq 20000$ kg/ms^2 ($\rho v^2 \geq 13400$ lb/fts^2)
	Gas	$\rho v^2 < 50000$ kg/ms^2 ($\rho v^2 < 33500$ lb/fts^2) Or $\rho v^2 < \frac{10000}{\sqrt{\frac{\mu_{gas}}{1 \cdot 10^{-3}}}} \frac{kg}{ms^2}$ $(\mu_{gas}\ in\ Pa \cdot s)$ $\left(\rho v^2 < \frac{6700}{\sqrt{\mu_{gas}}} \frac{lb}{fts^2}\right)$ $(\mu_{gas}\ in\ cP)$	-	$\rho v^2 \geq 50000$ kg/ms^2 ($\rho v^2 \geq 33500$ lb/fts^2) Or $\rho v^2 \geq \frac{10000}{\sqrt{\frac{\mu_{gas}}{1 \cdot 10^{-3}}}} \frac{kg}{ms^2}$ $(\mu_{gas}\ in\ Pa \cdot s)$ $\left(\rho v^2 \geq \frac{6700}{\sqrt{\mu_{gas}}} \frac{lb}{fts^2}\right)$ $(\mu_{gas}\ in\ cP)$

Where:

μ_{gas} gas is the dynamic viscosity in Pa.s (cP), p is density in kg/m^3 (lb/ft^3) and v is velocity in m/s (ft/s).

Piping system falls under medium and high susceptibility failure categories shall be further assessed and designed per Section 3, “Energy Institute guidelines for the avoidance of vibration induced fatigue failure in process pipework”.
If the piping system cannot meet detailed assessment “Energy Institute guidelines for the avoidance of vibration induced fatigue failure in process pipework” requirements for medium and high susceptibility failure categories, the piping system shall be evaluated using dynamic analysis methods.

Pulsation and transient vibrations
Pulsation due to periodic flow-induced excitation, which is only relevant for gas lines containing dead legs, is defined as follows;

$$\rho v^2 < \frac{400}{\pi\left(D/1000\right)^2} \quad (D \text{ in } mm)$$

where

- D is inside the diameter of the dead leg branch.

The above criteria are not met for dead legs, those dead legs in piping systems shall be assessed and designed per “Energy Institute guidelines for the avoidance of vibration induced fatigue failure in process pipework”.

If there are piping systems that will experience slug flow, or transient vibrations (flashing, cavitation, and surge/momentum) then an analysis as per the “Energy Institute guidelines for the avoidance of vibration induced fatigue failure in process pipework” shall be conducted.
If the piping system cannot meet detailed assessment “Energy Institute guidelines for the avoidance of vibration induced fatigue failure in process pipework” requirements for medium and high susceptibility failure categories, the

piping system shall be evaluated using dynamic analysis methods.

High-frequency acoustic excitation

High-frequency acoustic excitation is more predominant in gas piping systems with pressure reducing devices. This type of vibration takes the form of local pipe wall flexure resulting in potentially high dynamic stress levels at circumferential discontinuities on the pipe wall, fabricated tees or welded pipe supports, small-bore connections.

Effect of "high-frequency acoustic excitation" on piping systems, the sound power level of the piping system can be calculated as follows;

$$PWL = 10\log_{10}[\left(\frac{P_1-P_2}{P_1}\right)^{3.6} W^2 \left(\frac{T}{Mw}\right)^{1.2}] + 126.1 + SFF$$

SFF is a correction factor to account for multiple occurrences of sonic flow in a line. If consecutive sonic conditions exist, then SFF=6; otherwise SFF = 0.

Where:

- P_1 is upstream pressure (bara)
- P_2 is downstream pressure (bara)
- W is flow rate (kg/s)
- T is the upstream temperature (K)
- Mw is molecular weight (grams/mol)

Piping systems with sound power level (PWL) > 155 dB shall be further assessed and designed in accordance with "Quantitative" methods of Section T-2, Energy Institute guidelines for the avoidance of vibration induced fatigue failure in process pipework.

Transient/Surge analysis

Surge is caused by rapid velocity changes of the fluid, e.g. because of rapid valve closure, pump trip, or pump start. Pressure surge calculations are relevant to pipe handling incompressible fluids (liquids).

The magnitude of the pressure surge is determined by the velocity of the fluid in the pipe and the rate at which this velocity is changed. As a thumb rule, a sudden velocity change of 1 m/s (3.3 ft/s) will generate a pressure surge of 10 bar. The effect of a sudden velocity change depends on the length of the pipe and of the closing speed of the end valve.

The maximum surge pressure develops when the valve closes rapidly relative to the piping length. As per the ASME B31.3, the maximum permissible surge pressure shall be 1.33 times the system/piping design pressure. Measures should be taken to avoid cavitation since the collapse of a cavity could generate a high-pressure surge.

If the inlet pressure of a pipeline/piping system is reduced, e.g. due to a pump trip or due to closure of an upstream valve, a negative pressure wave travels through the system. The local pressure could drop below the vapor pressure of the fluid and a vapor cavity will form locally in the pipeline. The location of the cavity depends on the elevation profile of the pipe and will most likely be directly downstream of the equipment causing the flow reduction (e.g., pump or valve) and at high points in the system.

Surge pressure in piping systems shall be calculated as follows;

$$P_{Surge} = \frac{0.02 \times L}{D} \times \rho \times v_{fluid} \times v_{eff} \text{ when } \frac{L}{DN} \times v_{eff} < 0.55$$

$$P_{Surge} = c \times \rho \times v_{fluid} \times 10^{-5} \text{ when } \frac{L}{DN} \times v_{eff} \geq 0.55$$

where:

- L : Piping length between pump and valve (m)
- DN : Nominal diameter pipe (mm)
- V_{eff} : Effective valve closure speed (m/s)
- P_{surge} : Surge pressure (barg)

- V_{fluid} : Fluid velocity in pipe (m/s)
- ρ : Density of fluid (kg/m^3)

Piping systems that fail to meet the above criteria shall be subject to a detailed surge analysis, including evaluation of the following surge scenarios:

- Rapid valve closure with and without pump tripping;
- Pump(s) tripping;
- Pump start-up;
- Pump(s) trip followed by a restart.
- Pump kickback.

In general, Surge analysis shall be performed and documented for the following services:

- LNG loading and rundown pipes.
- LPG loading and rundown pipes.
- All hydrocarbon condensate loading pipes.
- Natural gas-liquid systems.
- The main distribution headers for cooling water systems.
- The main distribution headers for fire water systems.
- Piping running to/from tank car, rail car, barge, and ship loading arms and loading facilities.

Measures for mitigating surge pressure in pipeline

- Stronger pipework to withstand the pressure surge
- Rerouting piping
- Additional pipe supports
- Change of pipe material to one with a lower modulus (i.e. thermoplastic pipe materials)
- Flow control valve
- Air/Vacuum Release valves
- Intermediate check valves

- Non-slam check valves
- Bypass Valves
- Gas accumulators
- Liquid accumulators
- Surge tanks
- Surge shafts
- Surge anticipation valves
- Relief valves
- Bursting discs
- Weak pipe sections
- Increase diameter of pipeline to reduce average velocity
- Variable speed drives
- Soft starters
- Valve closure and opening times
- Increasing the inertia of pumps and motors (i.e. flywheels or by selection)

Simulation

Let's start with an example of a centrifugal pump suction line. Water flowing at 30 m^3 /h at 15 ° C and 0.3 barg pressure upstream pressure.

First select component Water and Fluid package IAPWS (International Association for the Properties of Water and Steam) in Aspen HYSYS.

Figure 5.1: Component list

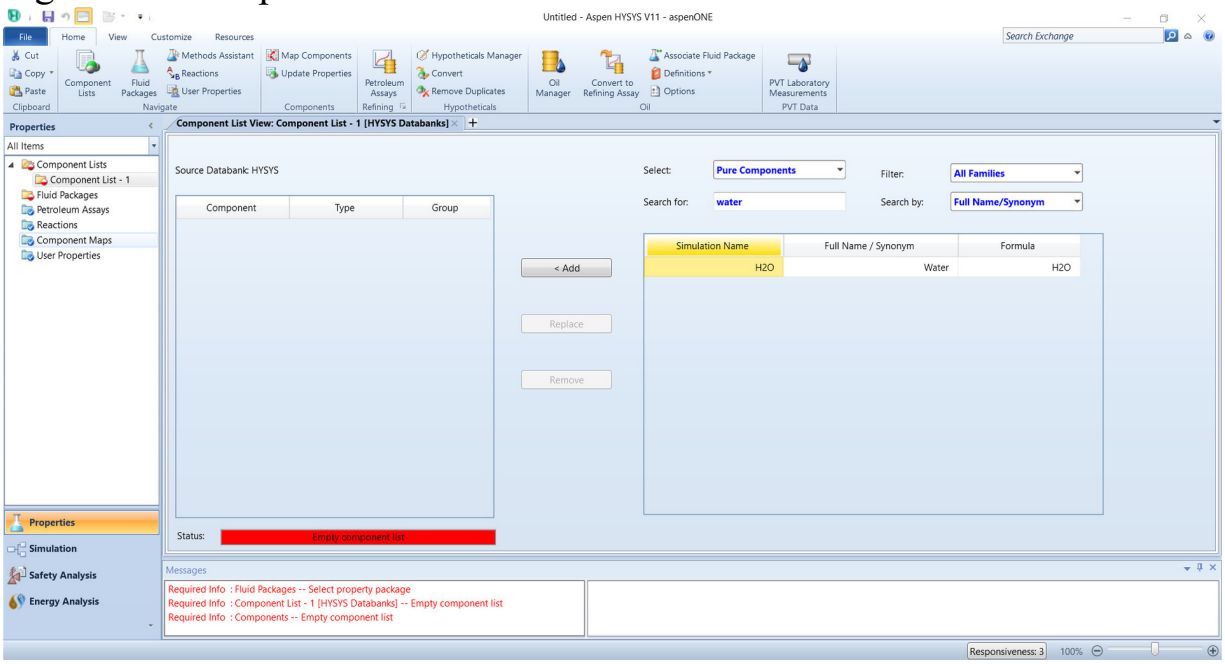
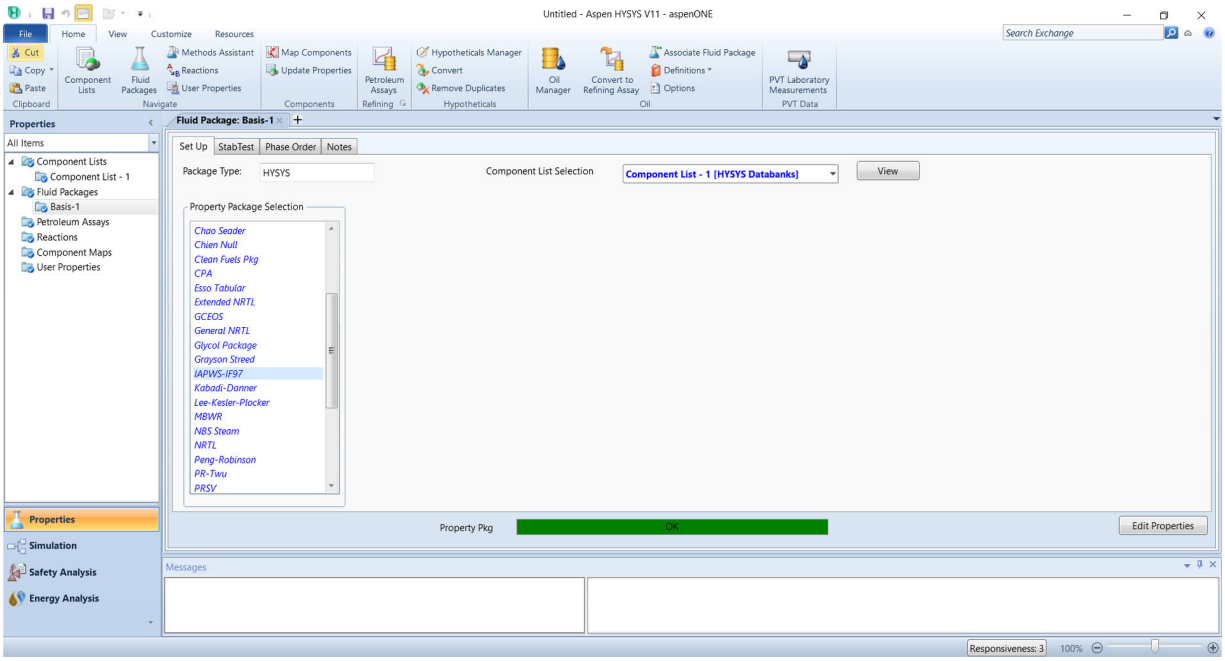
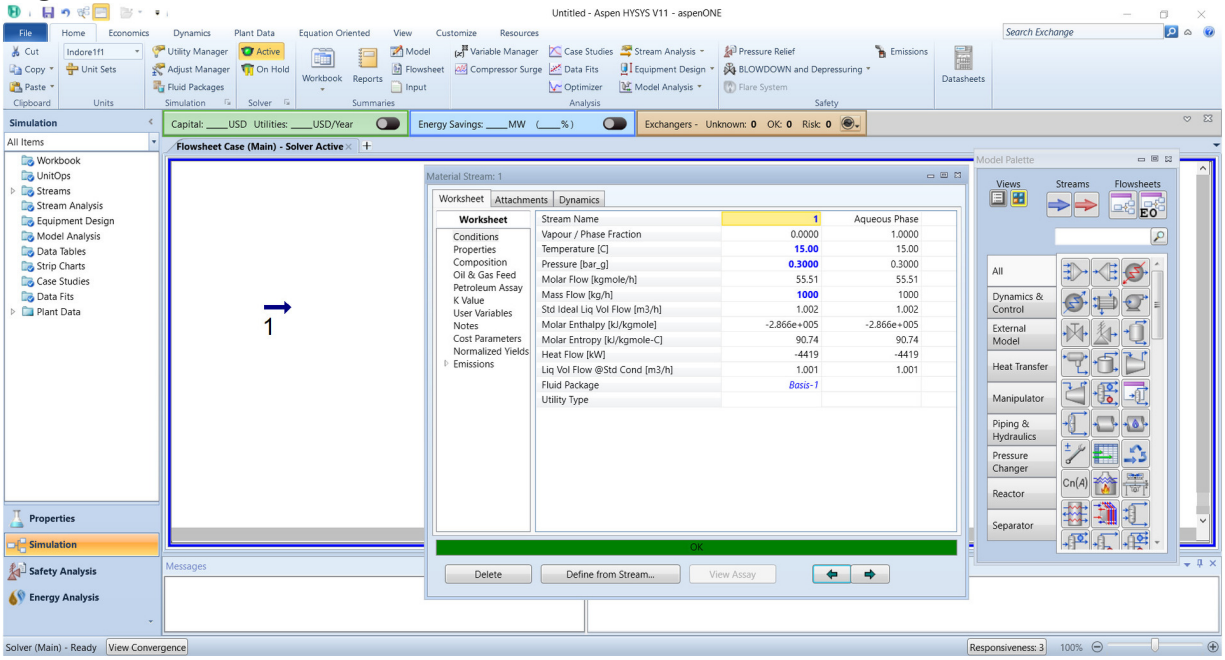


Figure 5.2: Fluid package


Untitled - Aspen HYSYS V11 - aspenONE
File
Home
View
Customize
Resources
Search Exchange
Cut
Copy
Paste
Clipboard
Component Lists
Fluid Packages
Methods Assistant
Reactions
User Properties
Navigate
Map Components
Update Properties
Components
Petroleum Assays
Refining
Hypotheticals Manager
Convert
Remove Duplicates
Hypotheticals
Oil Manager
Convert to Refining Assay
Associate Fluid Package
Definitions
Options
Oil
PVT Laboratory Measurements
PVT Data
Properties
All Items
Component Lists
Component List - 1
Fluid Packages
Basis-1
Petroleum Assays
Reactions
Component Maps
User Properties
Fluid Package: Basis-1
Set Up
StabTest
Phase Order
Notes
Package Type:
HYSYS
Component List Selection
Component List - 1 [HYSYS Databanks]
View
Property Package Selection
Chao Seader
Chien Null
Clean Fuels Pkg
CPA
Esso Tabular
Extended NRTL
GCEOS
General NRTL
Glycol Package
Grayson Streed
IAPWS-IF97
Kabadi-Danner
Lee-Kesler-Plocker
MBWR
NBS Steam
NRTL
Peng-Robinson
PR-Twu
PRSV
Property Pkg
OK
Edit Properties
Properties
Simulation
Safety Analysis
Energy Analysis
Messages
Responsiveness: 3
100%

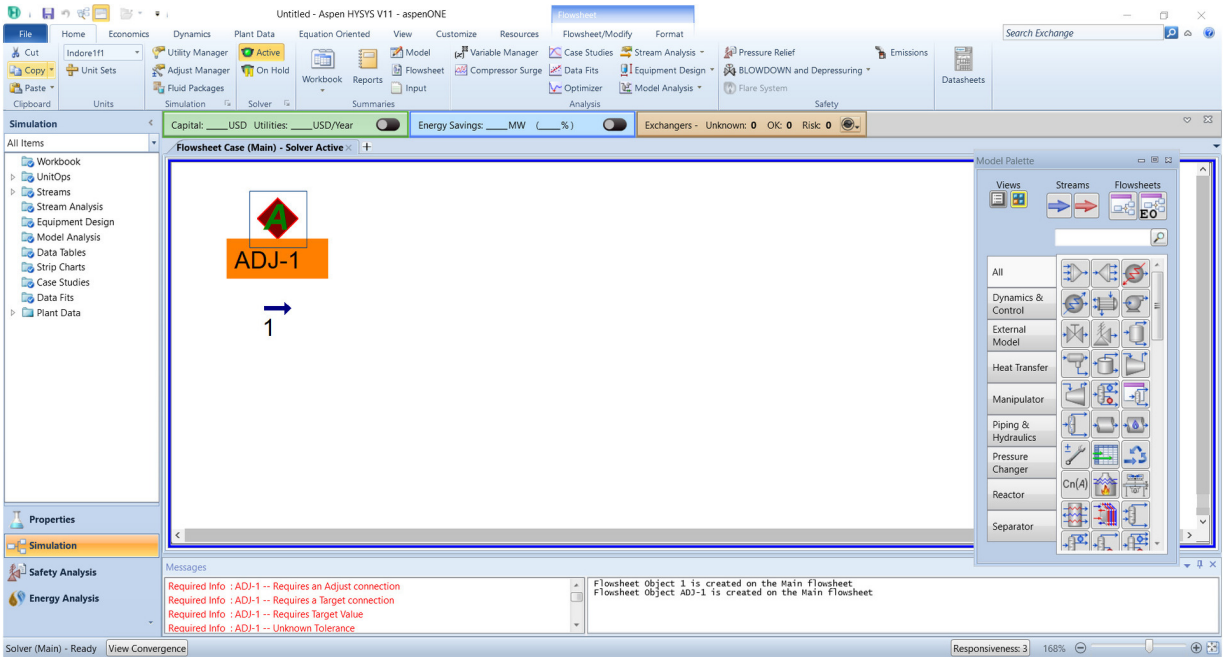
Use the model palette and model a material stream. Provide 15 ° C, 0.3 barg, 1000 kg/h mass flow rate and mole fraction 1. This will get the material stream converged.

Figure 5.3: Material stream



We will learn how to use the "Adjust" function. As per the problem statement, the water flow rate is 30 m^3 /h, so that will be the target and the variable will be the mass flow rate value of stream "1".
Select "Adjust" function from the model palette.

Figure 5.4: Adjust function



Double click on "ADJ-1" to open the Adjust window. Select "Target Variable" and "Adjusted variable".

Figure 5.5: Adjust function

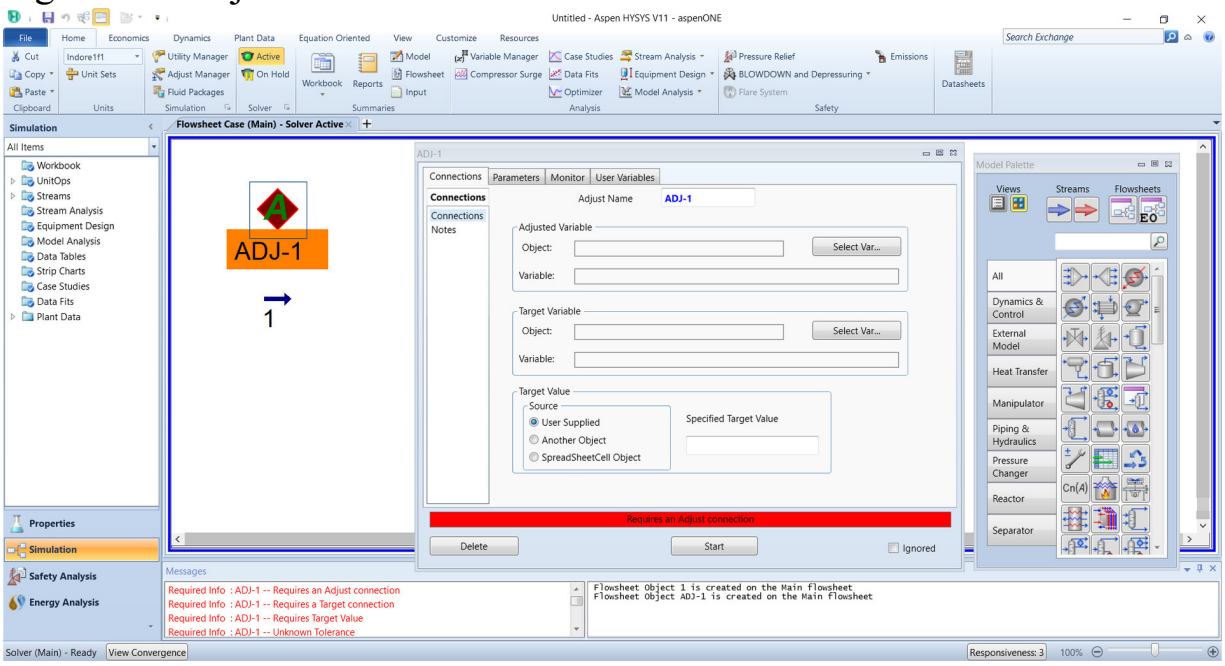


Figure 5.6: Adjust function

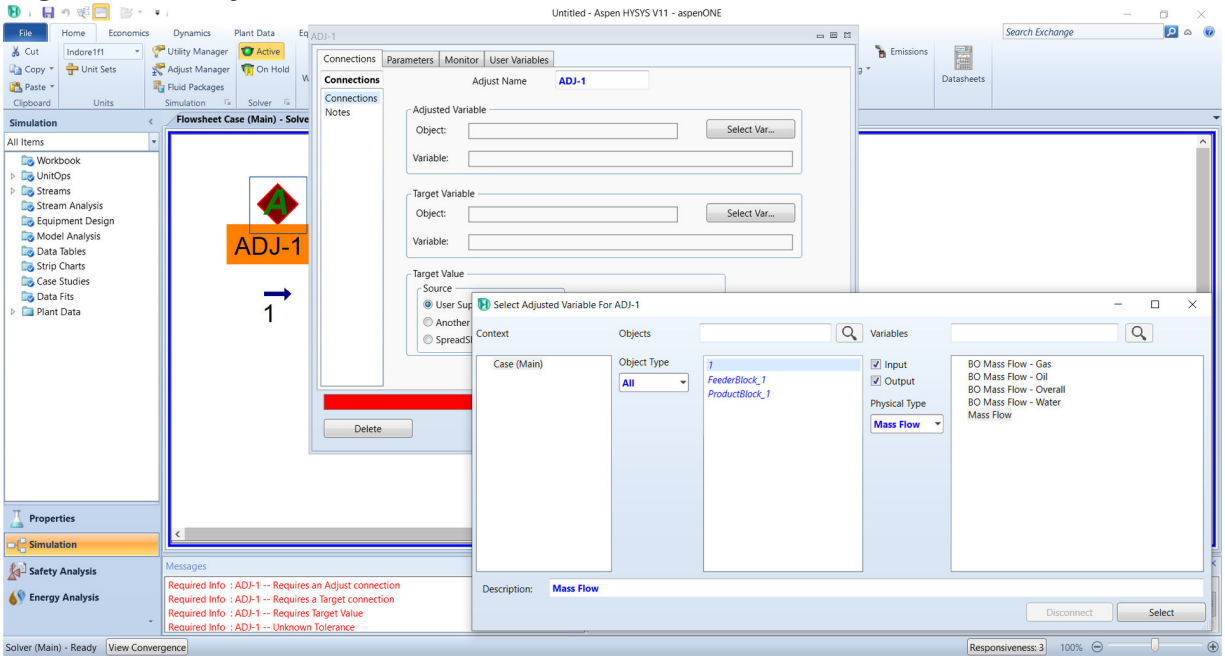


Figure 5.7: Adjust function

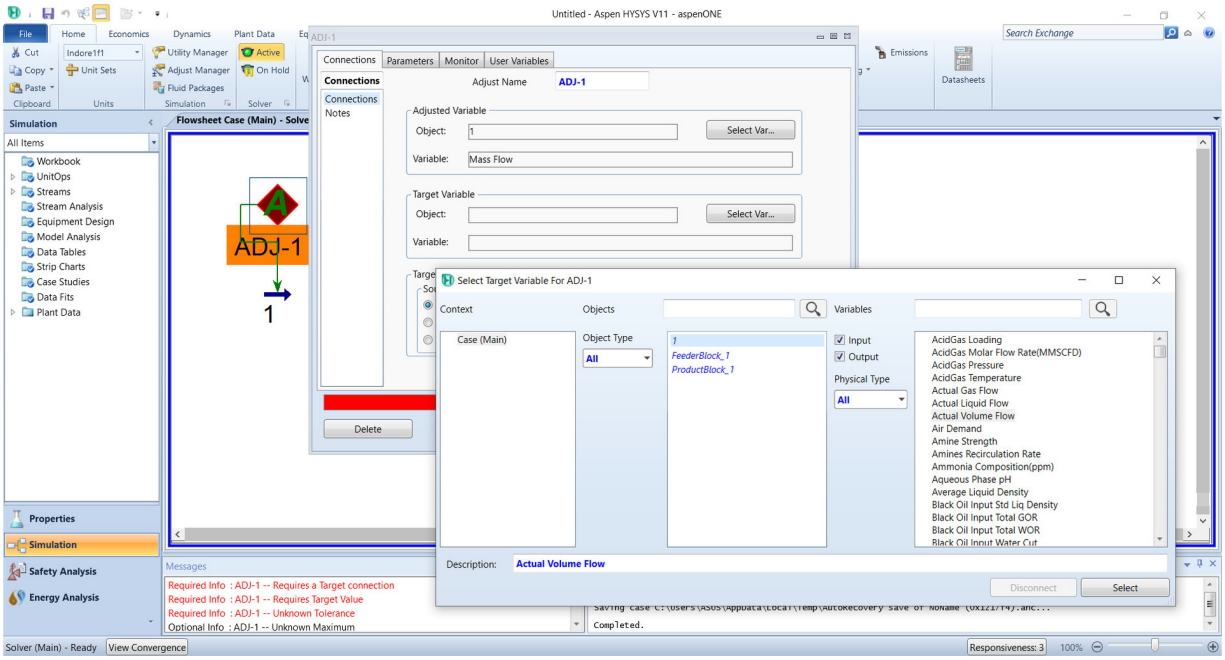
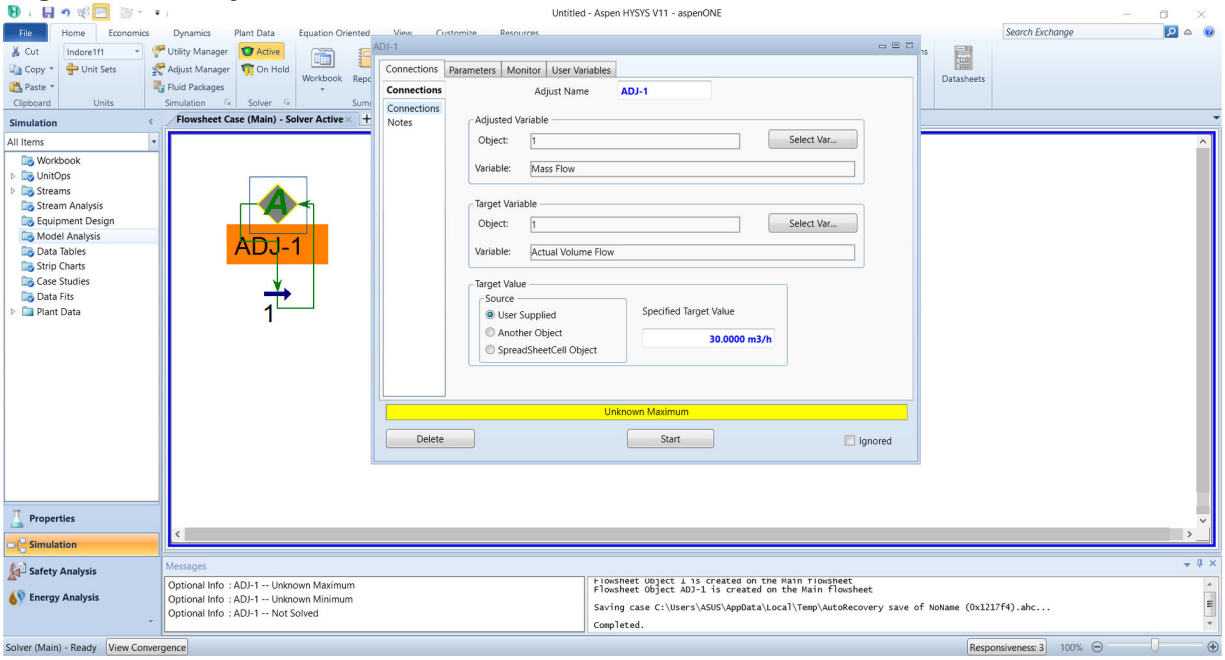
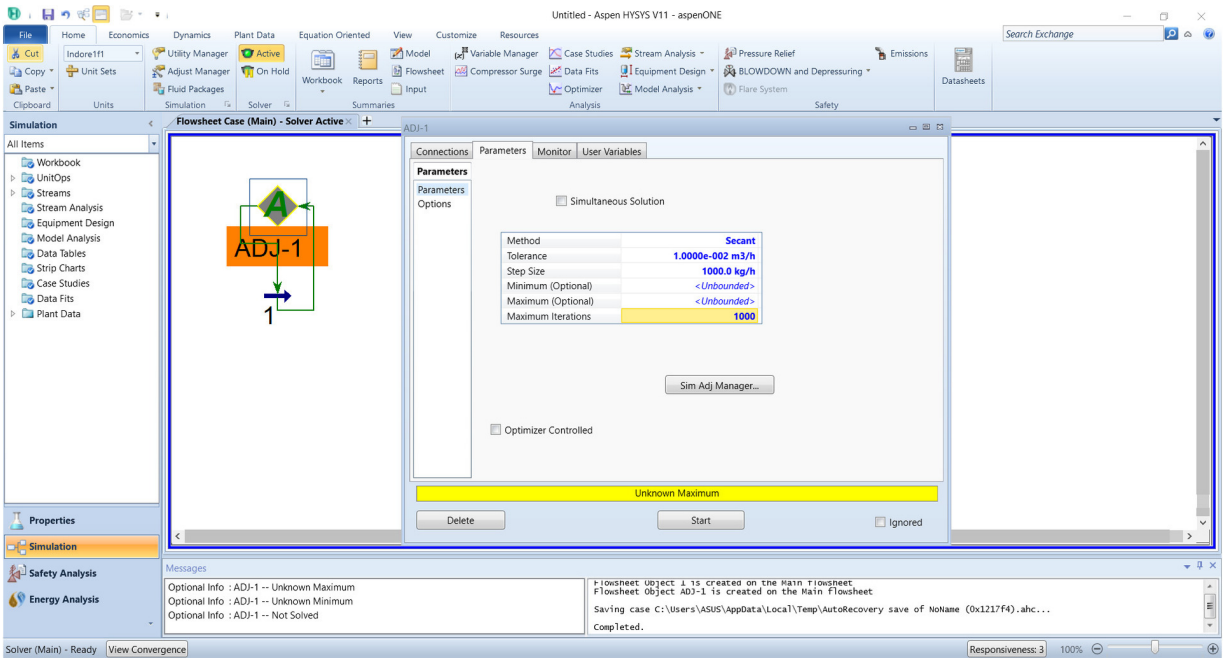


Figure 5.8: Adjust function



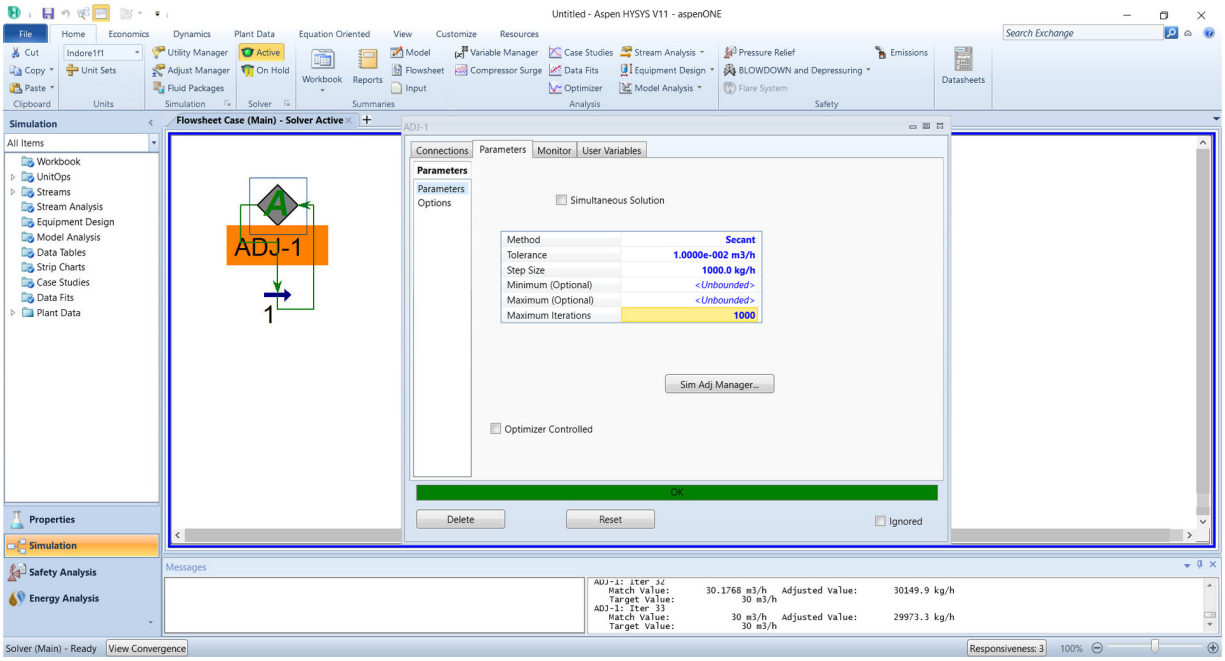
Next, select the “Parameters” tab and provide adequate tolerance, step size, and max. iterations values as shown below. Click “Start” to run the “Adjust” function. It will do the trial & error w.r.to mass flow rate and calculate the corresponding mass flow rate value to give 30 m^3 /h volume flow rate.

Figure 5.9: Adjust function



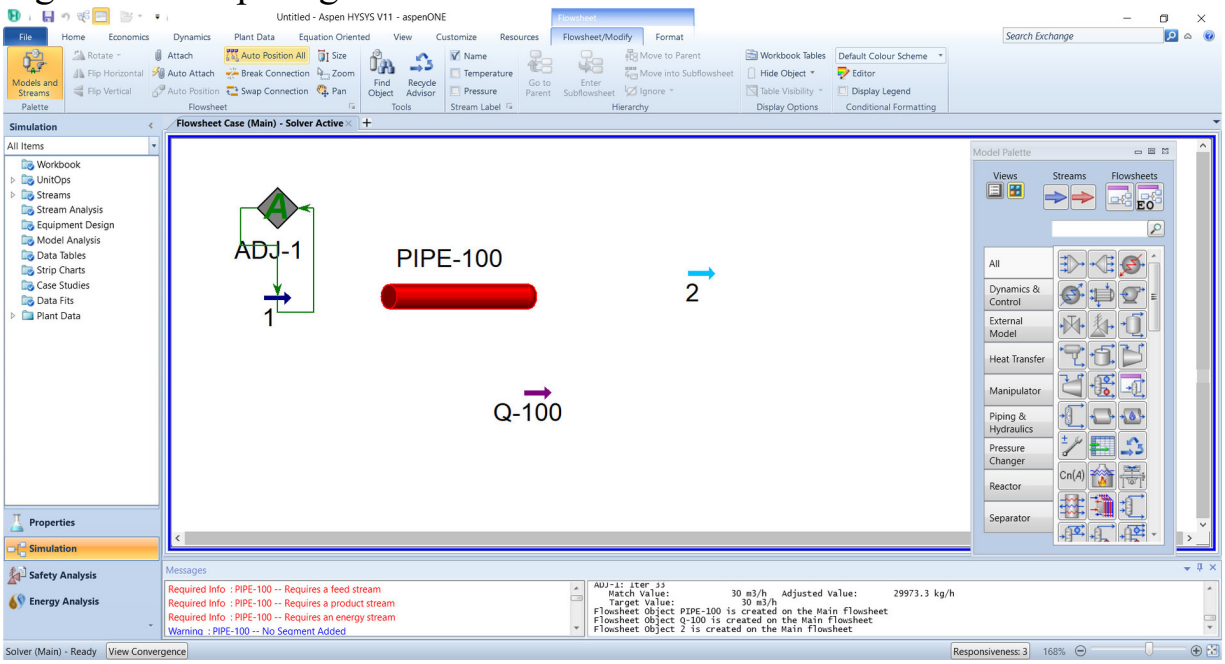
The back-calculated mass flow rate value is 29973.3 kg/h which is equal to 30 m^3 /h.

Figure 5.10: Adjust function


Untitled - Aspen HYSYS V11 - aspenONE
File
Home
Economics
Dynamics
Plant Data
Equation Oriented
View
Customize
Resources
Search Exchange
Cut
Copy
Paste
Clipboard
Indore1f1
Unit Sets
Units
Utility Manager
Adjust Manager
Fluid Packages
Simulation
Active
On Hold
Solver
Workbook
Reports
Model
Flowsheet
Input
Summaries
Variable Manager
Compressor Surge
Case Studies
Data Fits
Optimizer
Stream Analysis
Equipment Design
Model Analysis
Analysis
Pressure Relief
BLOWDOWN and Depressuring
Flare System
Emissions
Safety
Datasheets
Simulation
All Items
Workbook
UnitOps
Streams
Stream Analysis
Equipment Design
Model Analysis
Data Tables
Strip Charts
Case Studies
Data Fits
Plant Data
Properties
Simulation
Safety Analysis
Energy Analysis
Flowsheet Case (Main) - Solver Active
ADJ-1
1
ADJ-1
Connections
Parameters
Monitor
User Variables
Parameters
Parameters
Options
Simultaneous Solution
Method
Secant
Tolerance
1.0000e-002 m3/h
Step Size
1000.0 kg/h
Minimum (Optional)
<Unbounded>
Maximum (Optional)
<Unbounded>
Maximum Iterations
1000
Sim Adj Manager...
Optimizer Controlled
OK
Delete
Reset
Ignored
Messages
ADJ-1: Iter 32
Match Value: 30.1768 m3/h Adjusted Value: 30149.9 kg/h
Target Value: 30 m3/h
ADJ-1: Iter 33
Match Value: 30 m3/h Adjusted Value: 29973.3 kg/h
Target Value: 30 m3/h
Solver (Main) - Ready
View Convergence
Responsiveness: 3
100%

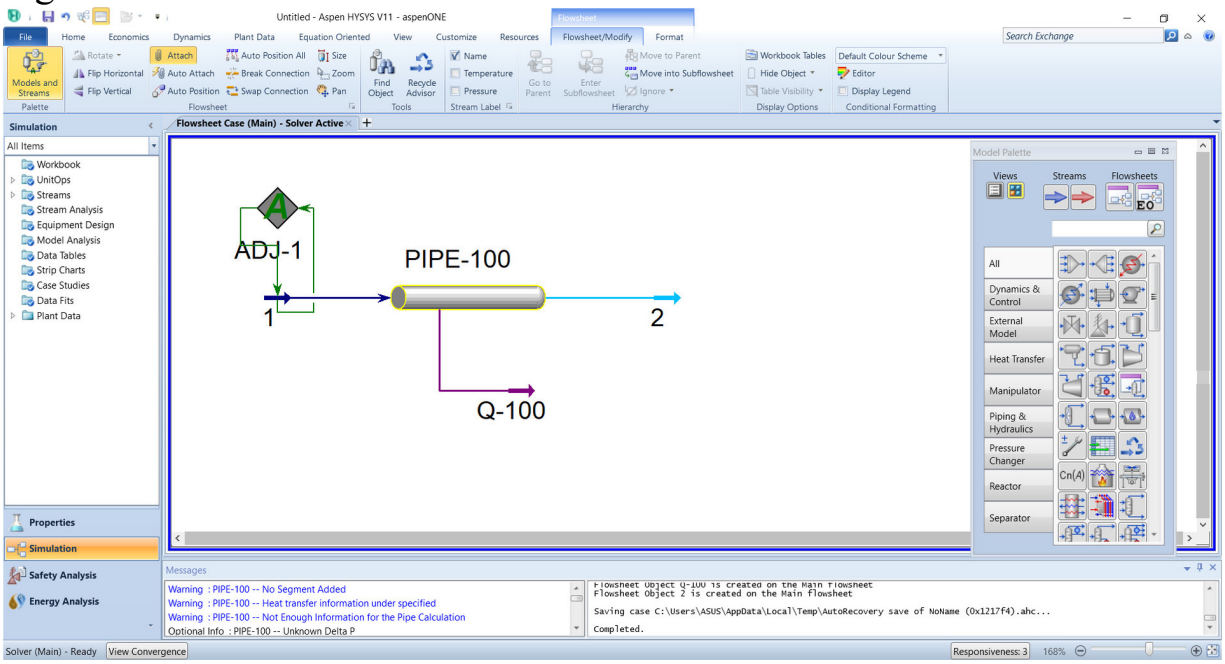
Use Model Palette to insert the Pipe segment, Material Stream and Energy Stream.

Figure 5.11: Pipe segment



Use the “Attach” function to connect the streams.

Figure 5.12: Attach function



We will consider no heat loss or gain for this pipe segment. Double click on energy stream and write heat flow value as zero.

Figure 5.13: Energy stream

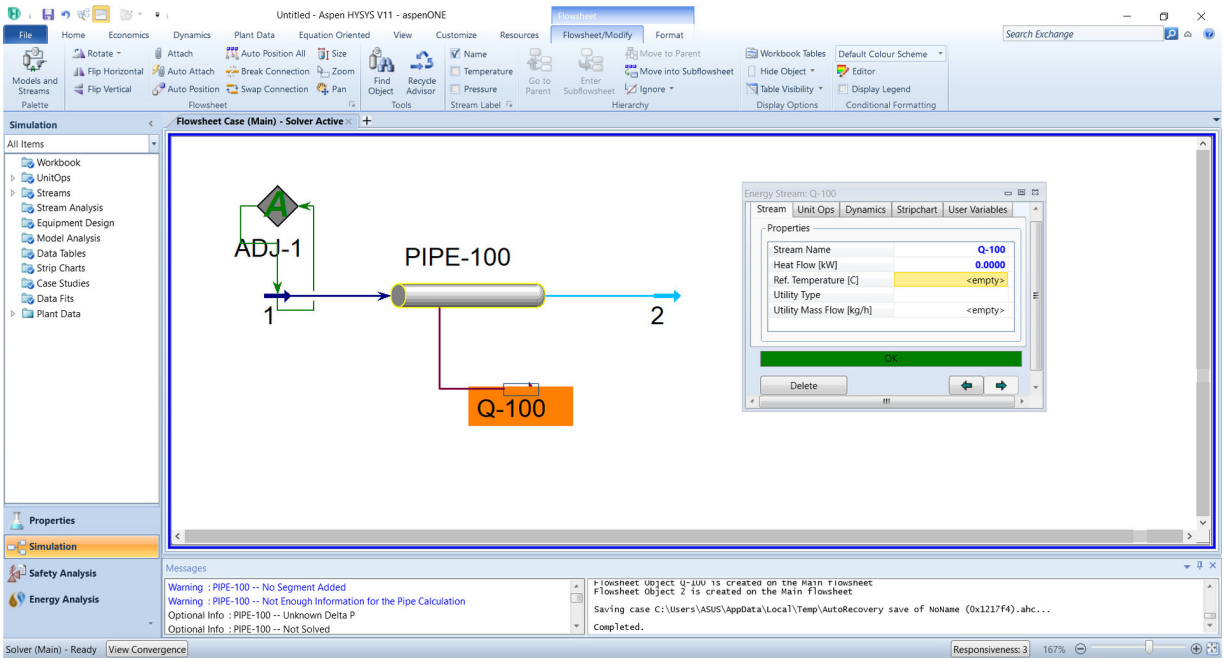
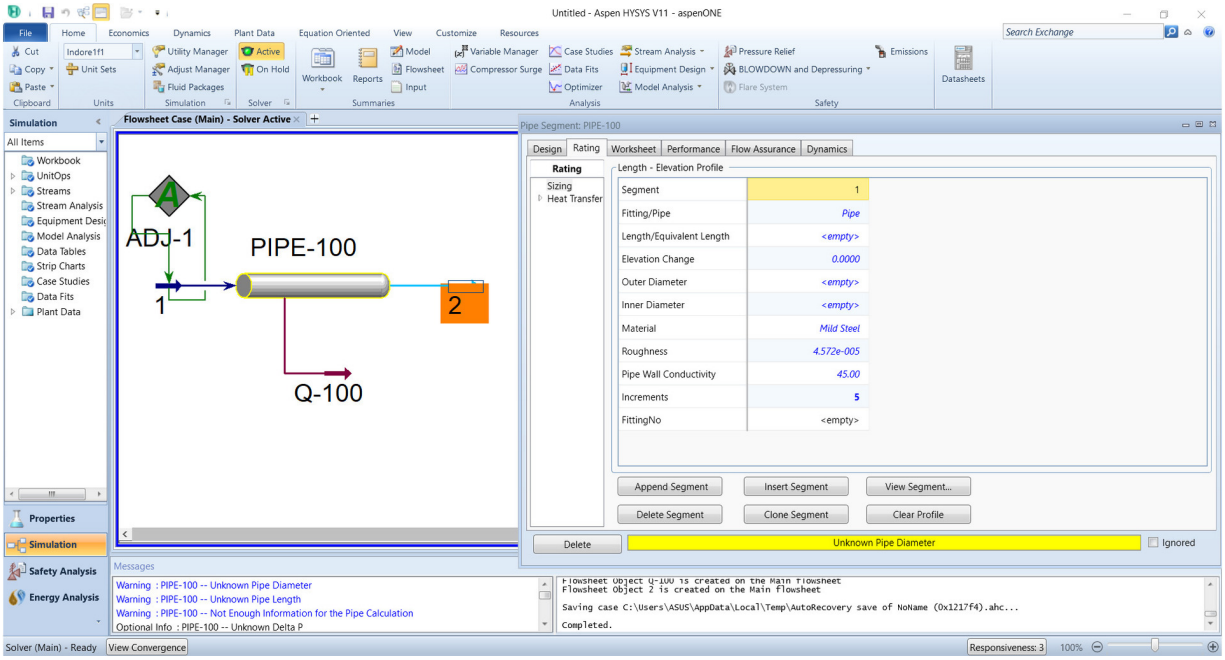
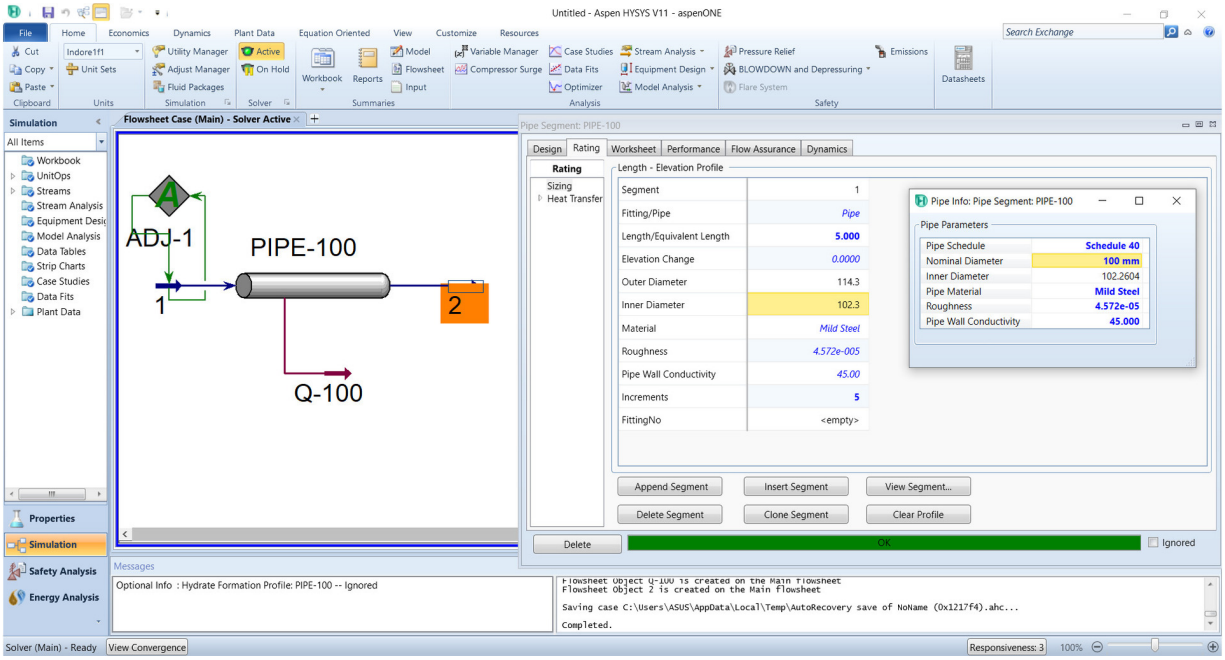


Figure 5.14: Pipe segment



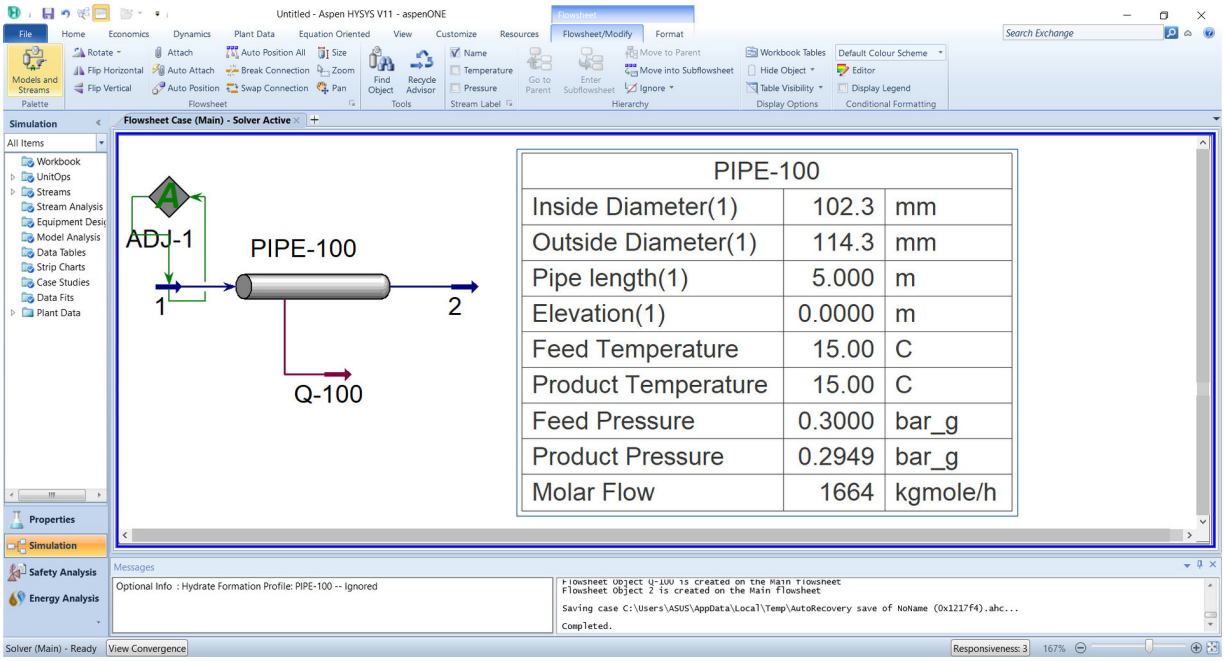
Provide 5 m as pipe length. Double click on Outer Diameter value to open a new window, wherein you can input pipe schedule, Nominal diameter, and pipe MOC data. One can add more segments as pipes or fittings using the same “Append Segment” tab.

Figure 5.15: Pipe segment



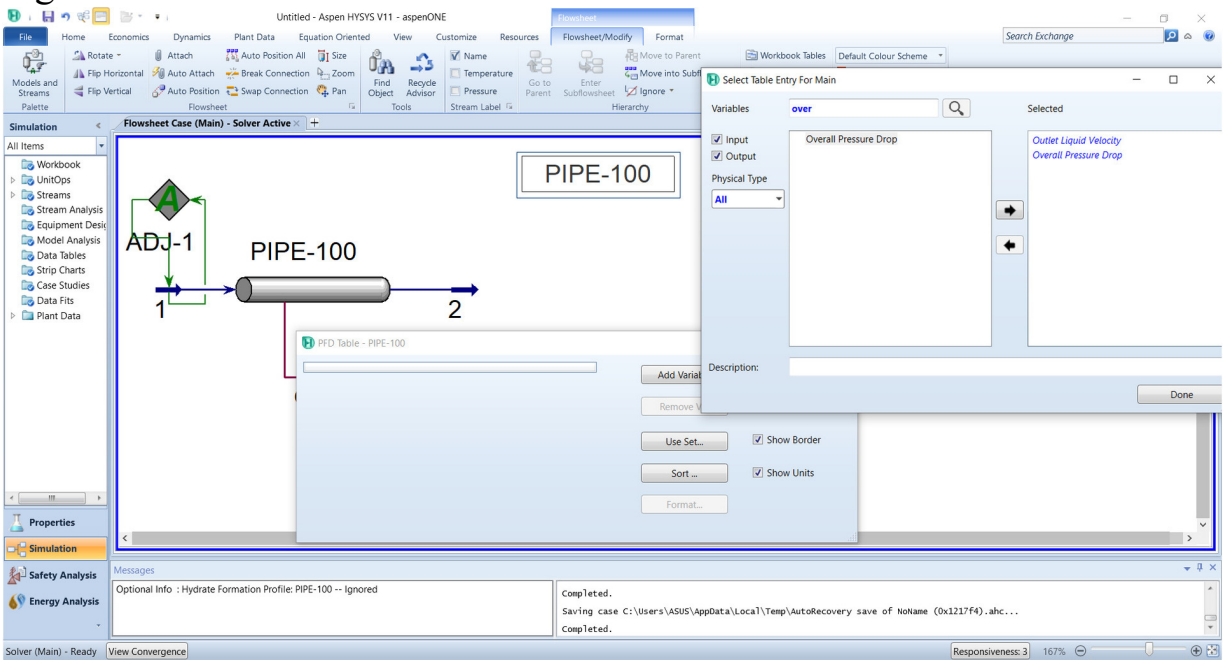
Right-click on Pipe segment (PIPE-100) and select “Show table”.

Figure 5.16: Show table


Untitled - Aspen HYSYS V11 - aspenONE
Flowsheet Case (Main) - Solver Active
ADJ-1
PIPE-100
1
2
Q-100
PIPE-100
Inside Diameter(1) 102.3 mm
Outside Diameter(1) 114.3 mm
Pipe length(1) 5.000 m
Elevation(1) 0.0000 m
Feed Temperature 15.00 C
Product Temperature 15.00 C
Feed Pressure 0.3000 bar_g
Product Pressure 0.2949 bar_g
Molar Flow 1664 kgmole/h
Optional Info : Hydrate Formation Profile: PIPE-100 -- Ignored
Flowsheet Object 2 is created on the Main Flowsheet
Saving case C:\Users\ASUS\AppData\Local\Temp\AutoRecovery save of NoName (0x1217f4).ahc...
Completed.
Solver (Main) - Ready
View Convergence

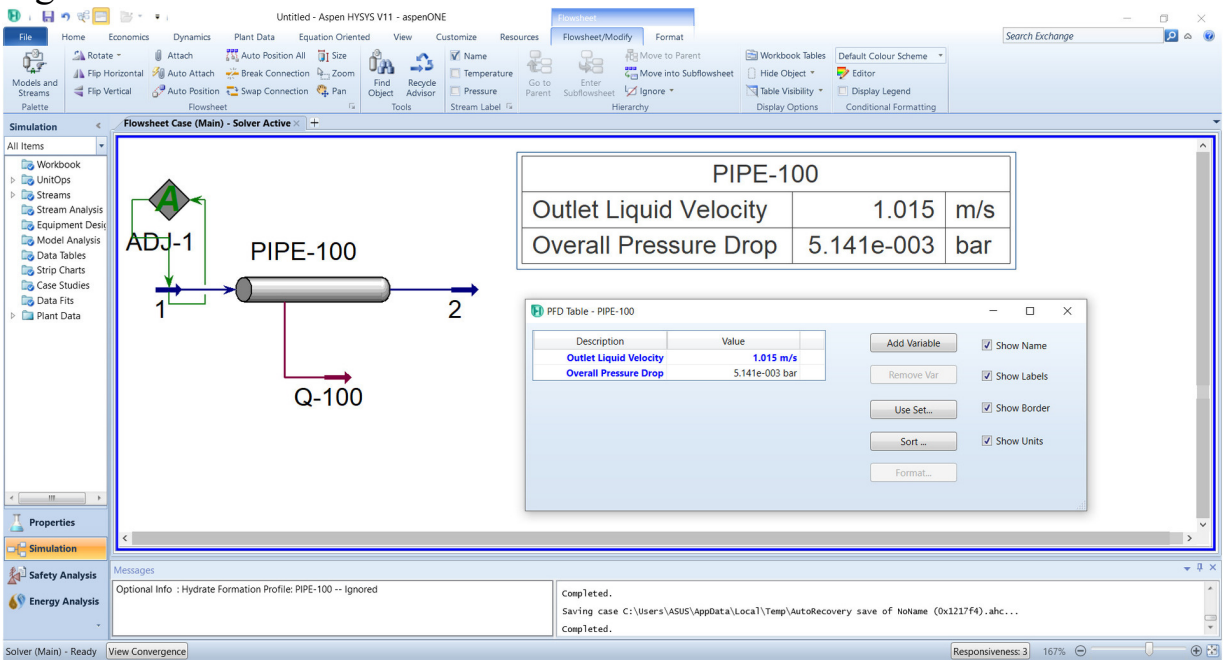
Double click on the table.

Figure 5.17: Show table



It may be noted that pump suction velocity calculated is about 1 m/s and the pressure drop is 0.005 bar.

Figure 5.18: Show table



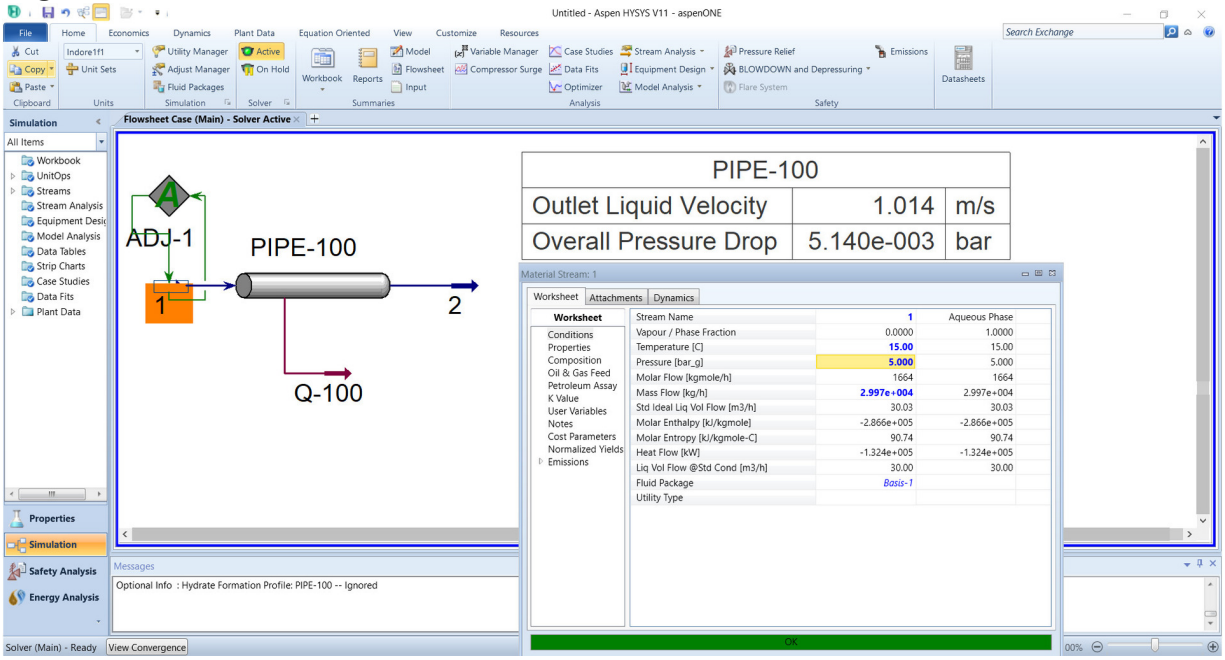
As per API 14E, for the centrifugal pump suction line, the typical velocity should be around 1 m/s. Hence selected 100 mm NB pipe is adequate. Velocity criteria must be satisfied and pressure drop criteria are nice to follow.

Table 5.2: Velocity criteria

	Suction Velocity (feet per second)	**Discharge Velocity (feet per second)**
Reciprocating Pumps		
Speeds up to 250 RPM	2	6
Speeds 251-330 RPM	1.5	4.5
Speeds above 330 RPM	1	3
Centrifugal Pumps	2-3	6-9

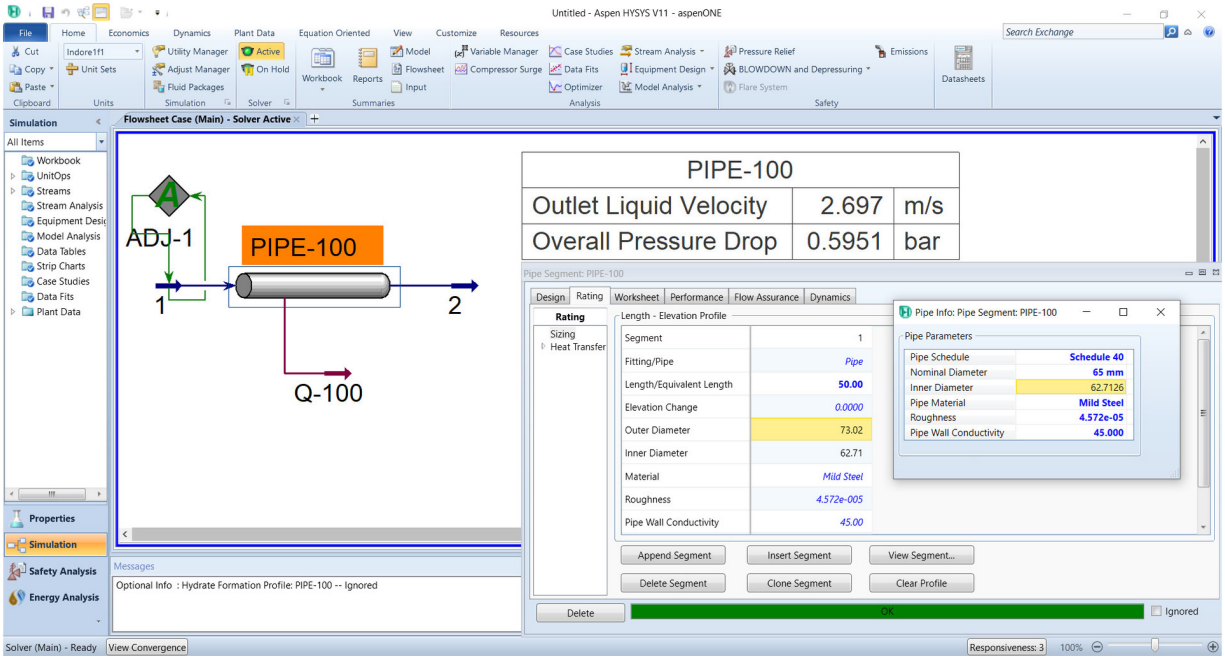
Now let's change the problem statement from centrifugal pump suction line to discharge line. Pipe length shall be 50 m. Our objective is to decide an appropriate line size.
Double click on Material Stream "1" and change pressure from 0.3 barg to 5 barg (as this is pump discharge point).

Figure 5.19: Material stream



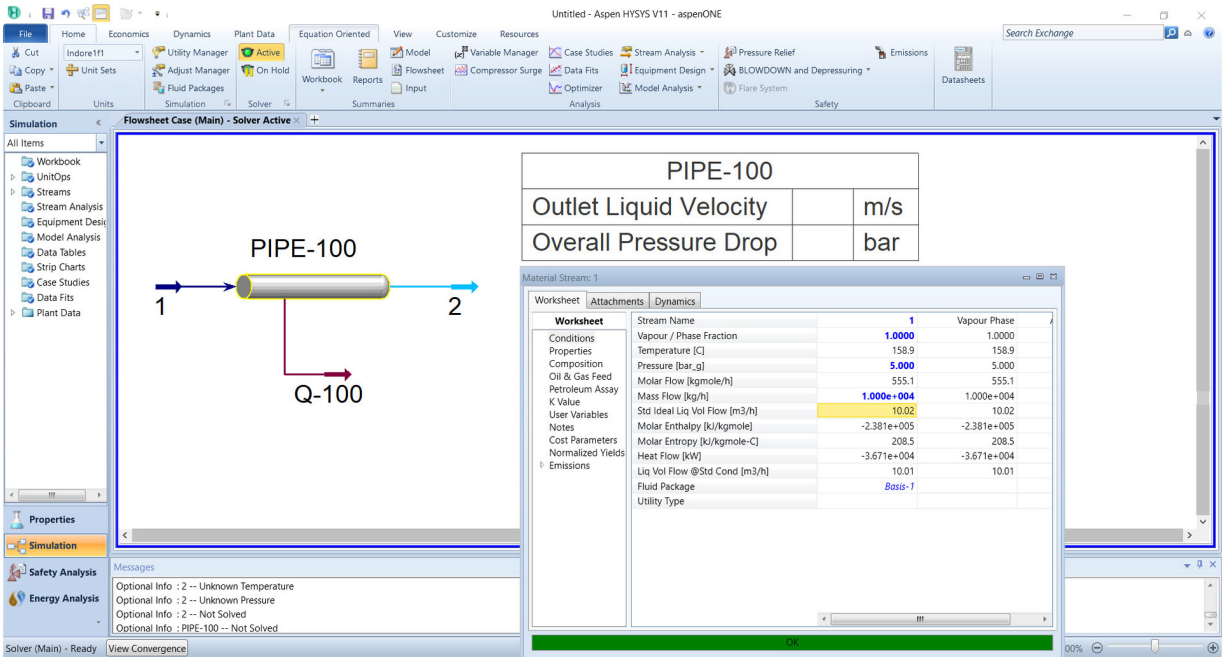
Double click on pipe segment, change the pipe length to 50 m, and change line size from 100 mm to 65 mm. The calculated pipe velocity is 2.7 m/s and the pressure drop is 0.6 bar. As per API 14E, the centrifugal pump discharge pipe typical velocity is around 2.7 m/s. Hence 65 mm NB pipe is adequate.

Figure 5.20: Pipe segment



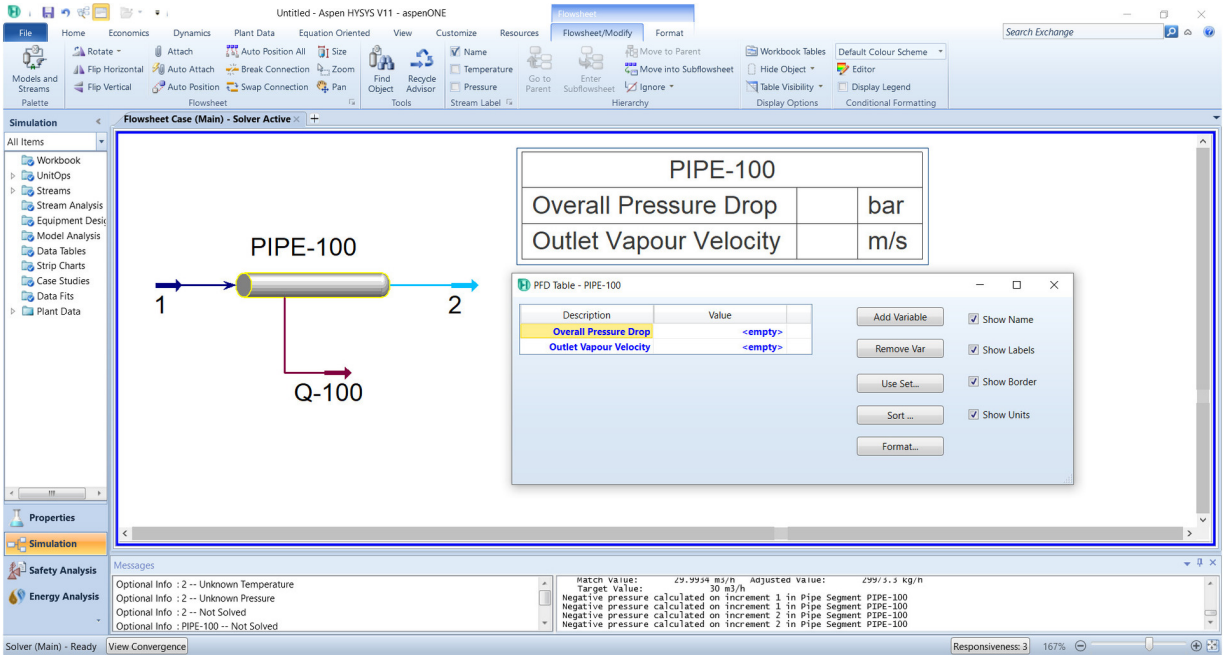
Now let's change the problem statement again. Instead of feed as liquid, we will change it to saturated steam at 5 barg pressure. Adjust function is deleted. The mass flow rate of steam is 10000 kg/h. Temperature is deleted. The vapor fraction is 1. Saturation temperature calculated by HYSYS is 158.9 ° C.

Figure 5.21: Material stream



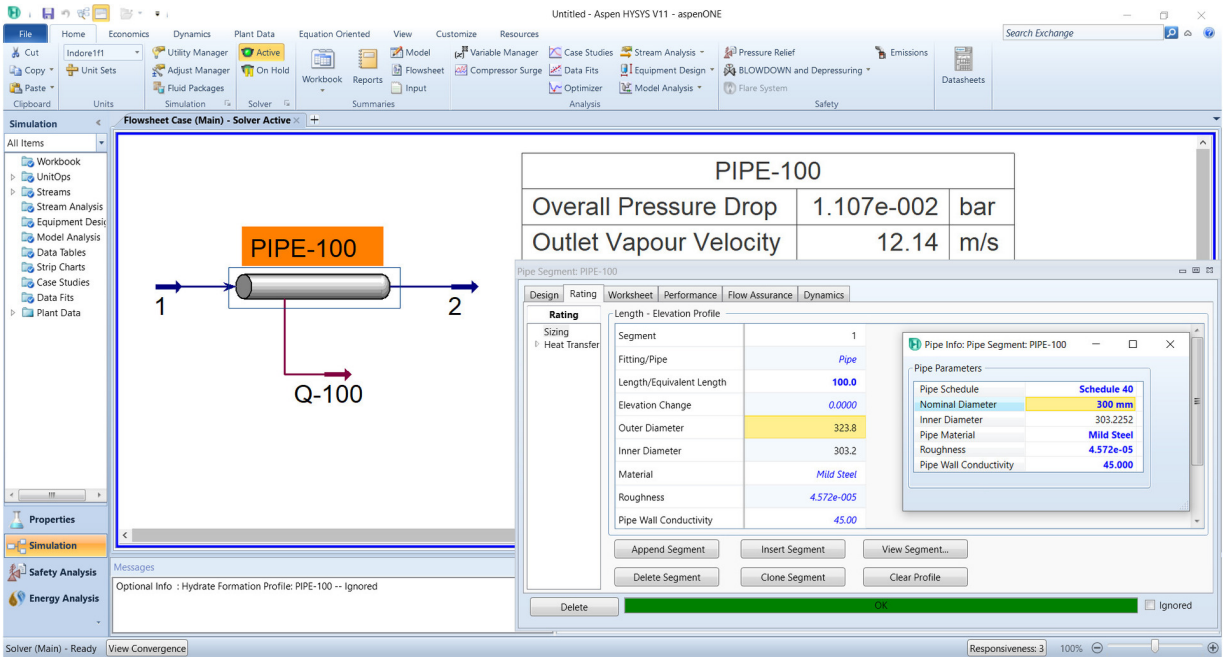
Double click on Pipe table. Replace liquid velocity by Outlet vapor velocity.

Figure 5.22: Show table



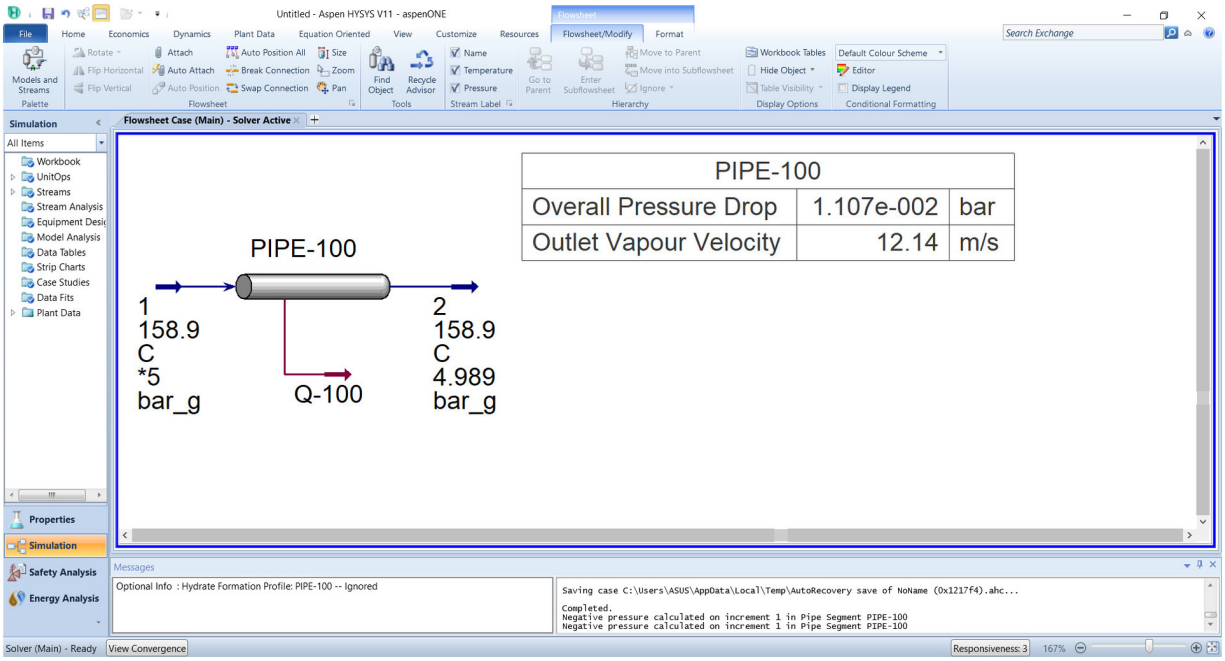
Double click on the pipe segment and change pipe NB from 65 NB to 300 NB. Pipe length shall be 100 m. The calculated velocity is 12 m/s and pressure drop 0.005 bar. As per API 14E gas maximum allowable velocity is 18 m/s. As above velocity is less than 18 m/s, 300 mm NB pipe size is adequate.

Figure 5.23: Pipe segment



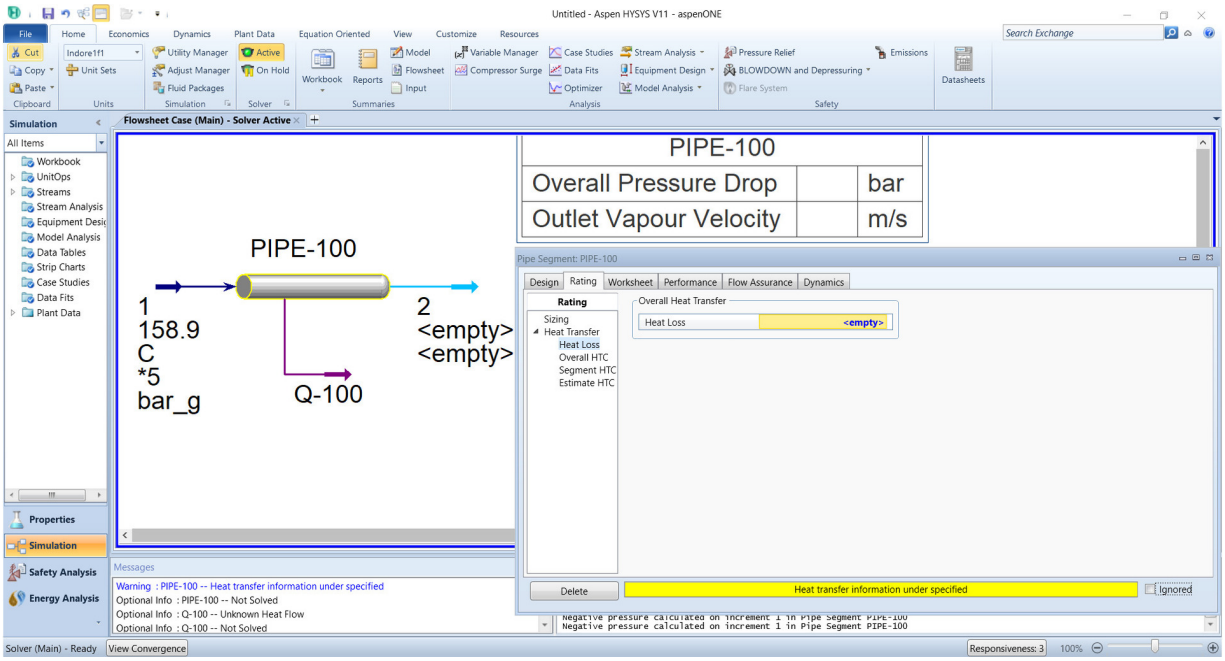
Select “Flowsheet/Modify” tab and check “Temperature” and “Pressure” boxes active in “Stream label”. This will make temperature and pressure values displayed on the flowsheet.

Figure 5.24: Pipe segment



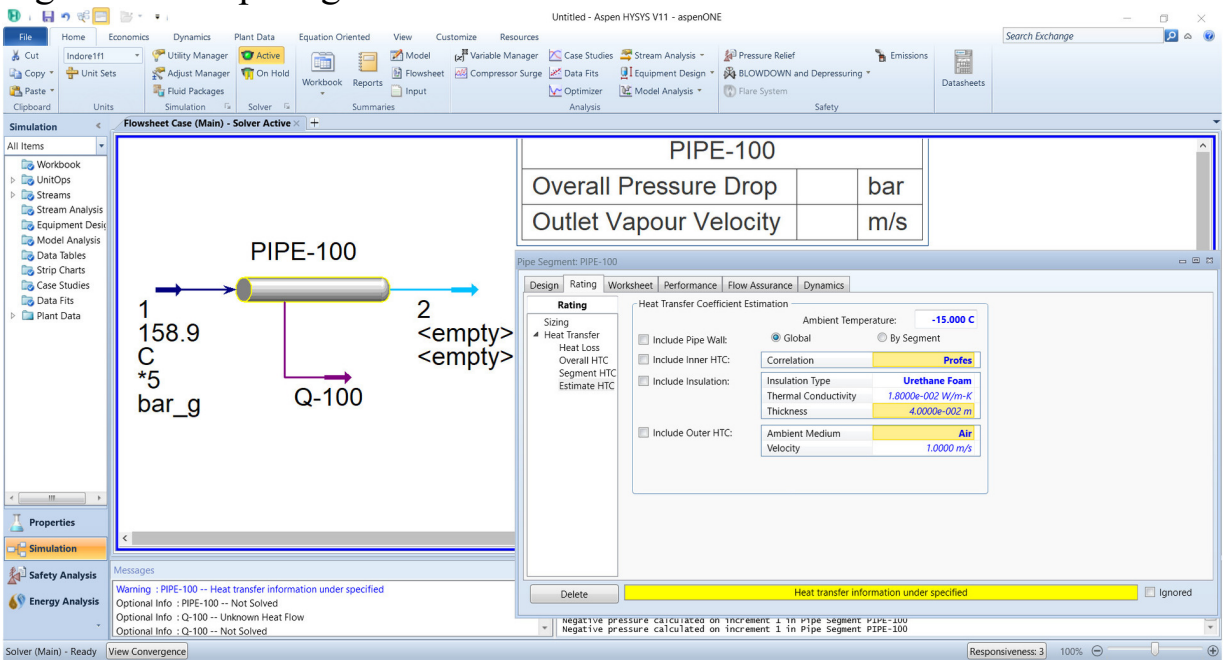
It may be noted that there is no temperature drop in this 100 m long pipe as we had given Q-100 value as zero. Let's make HYSYS calculate the heat loss. Select expand "Heat Transfer" tab and delete heat loss value (zero) in the "Heat Loss" tab. Then select "Estimate HTC". Heat transfer coefficient shall be calculated by HYSYS in this tab.

Figure 5.25: Pipe segment heat transfer



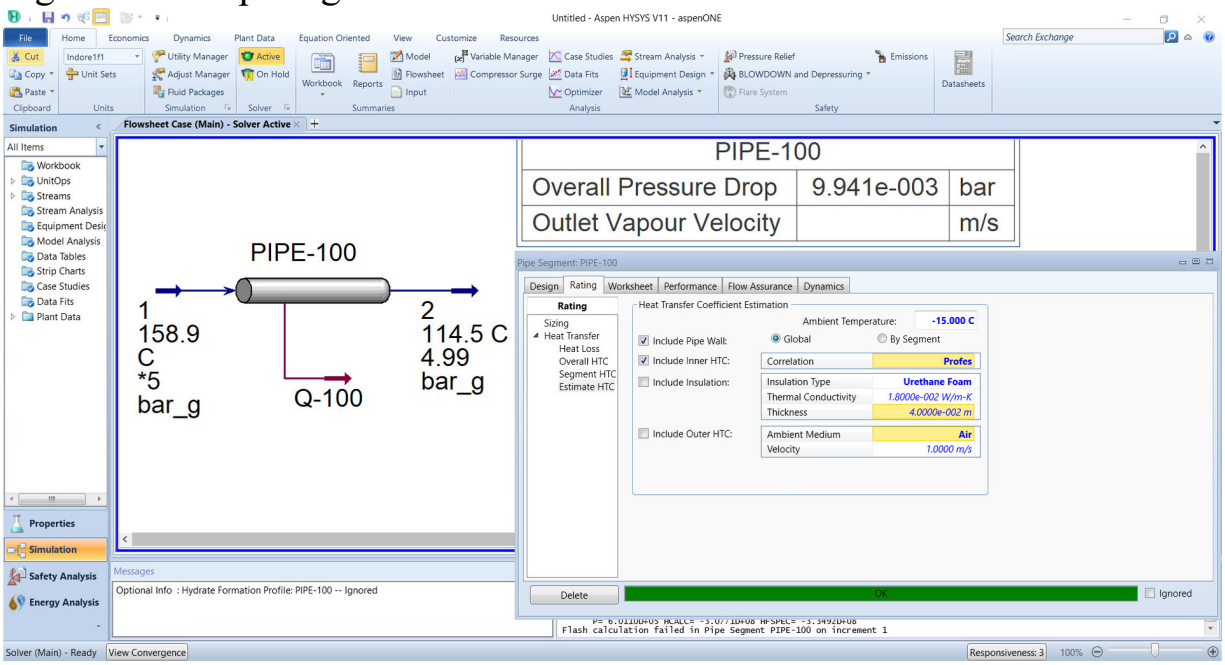
Provide ambient temperature, Insulation type and thickness, ambient medium, and air velocity as shown below.

Figure 5.26: Pipe segment heat transfer



Now checkboxes for “Include Pipe Wall” and “Inner Wall HTC”. It can be noted that a 100 m long pipe without any insulation would have considerable temperature drop. Hence, insulation is needed in this case.

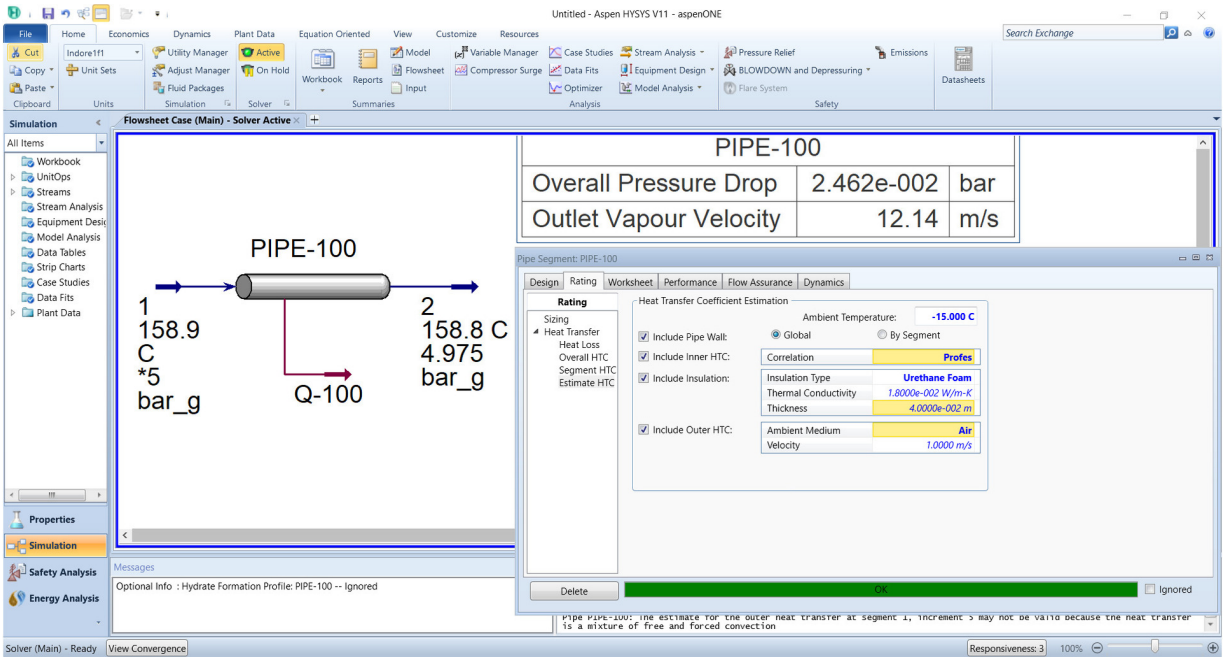
Figure 5.27: Pipe segment heat transfer



- If you check, 'include pipe wall" button it will estimate the conductive heat transfer based on the pipe material you specify.
- If you check, 'include inner HTC" button, it will estimate the convective heat transfer due to fluid movement.
- If you check, 'include insulation" button it will estimate the conductive heat transfer based on the insulation material you specify.
- If you check, 'include outer HTC" button it will estimate the **convective/conductive** heat transfer to the surrounding. --> Outer heat transfer

With the insulation box checked, there is a temperature drop on only 0.1 ° C.

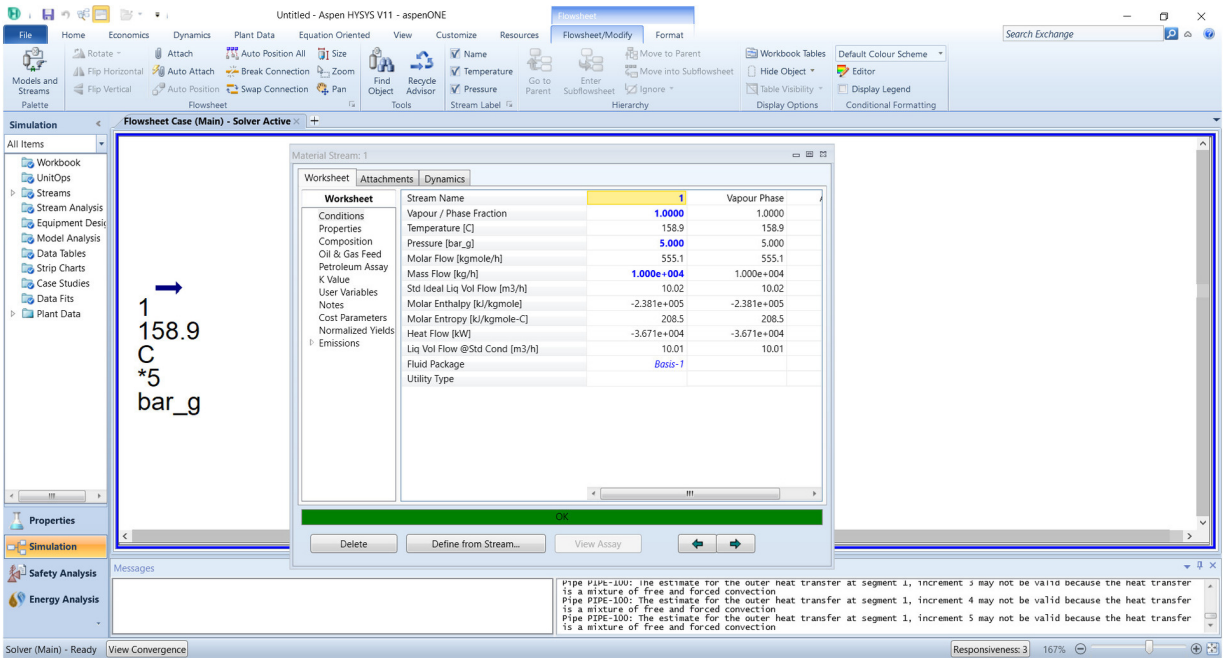
Figure 5.28: Pipe segment heat transfer



There is a quick way to carry out line sizing, that works well for single-phase fluids. We will choose the

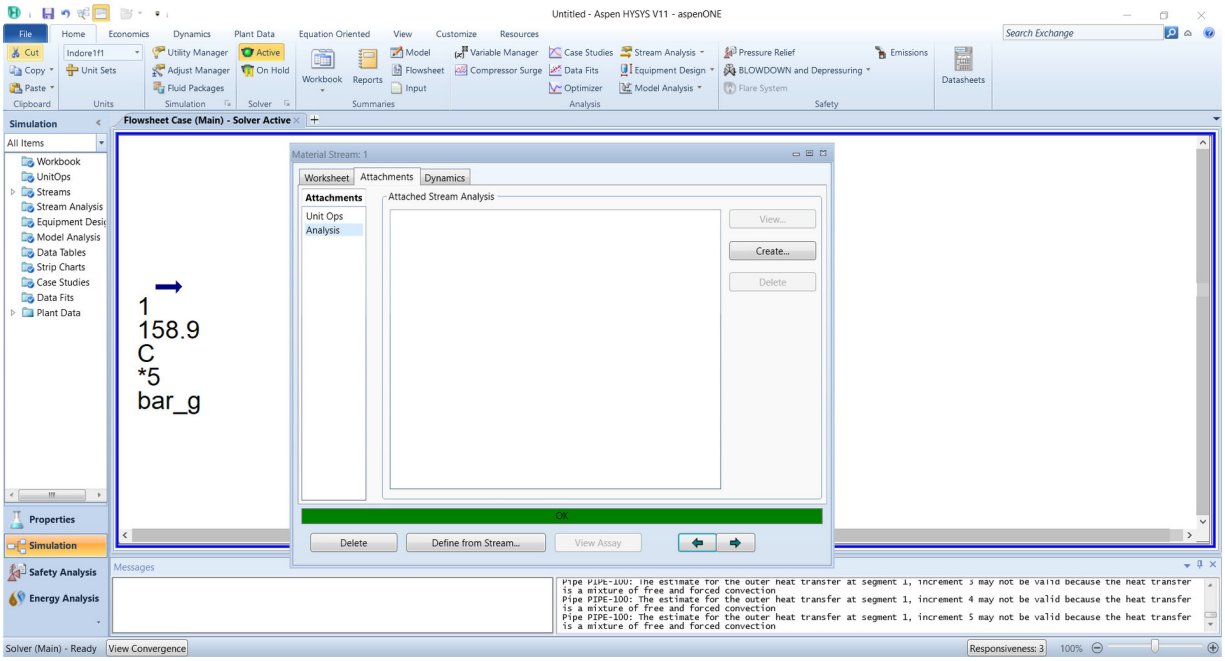
appropriate line size for stream “1”. This is the same material stream as the last example (vapor line sizing).

Figure 5.29: Material stream



Double click on a material stream and select the “Attachments\Analysis” tab.

Figure 5.30: Line sizing manager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Units
Utility Manager
Adjust Manager
Fluid Packages
Simulation
Active
On Hold
Solver
Workbook
Reports
Model
Flowsheet
Input
Summaries
Variable Manager
Compressor Surge
Case Studies
Data Fits
Optimizer
Stream Analysis
Equipment Design
Model Analysis
Analysis
Pressure Relief
BLOWDOWN and Depressuring
Flare System
Emissions
Safety
Datasheets
Simulation
All Items
Workbook
UnitOps
Streams
Stream Analysis
Equipment Desi
Model Analysis
Data Tables
Strip Charts
Case Studies
Data Fits
Plant Data
Flowsheet Case (Main) - Solver Active
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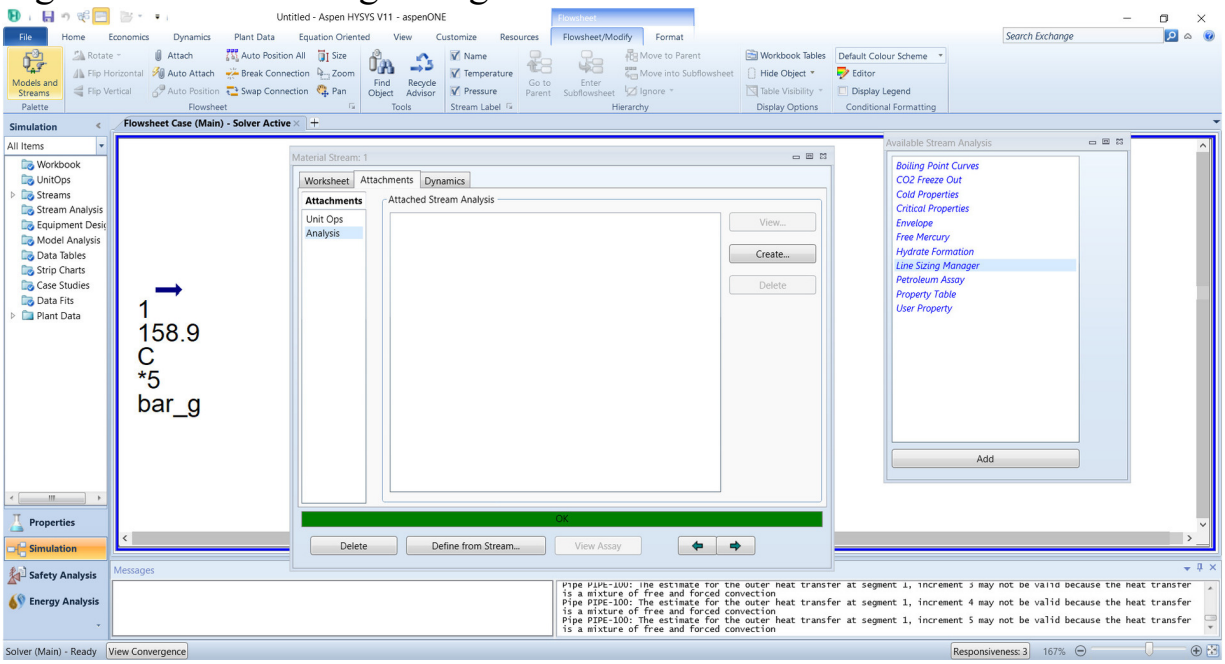
bar_g
Material Stream: 1
Worksheet
Attachments
Dynamics
Attachments
Unit Ops
Analysis
Attached Stream Analysis
View...
Create...
Delete
OK
Delete
Define from Stream...
View Assay
Properties
Simulation
Safety Analysis
Energy Analysis
Messages
Pipe PIPE-100: The estimate for the outer heat transfer at segment 1, increment 3 may not be valid because the heat transfer is a mixture of free and forced convection
Pipe PIPE-100: The estimate for the outer heat transfer at segment 1, increment 4 may not be valid because the heat transfer is a mixture of free and forced convection
Pipe PIPE-100: The estimate for the outer heat transfer at segment 1, increment 5 may not be valid because the heat transfer is a mixture of free and forced convection
Solver (Main) - Ready
View Convergence
Responsiveness: 3
100%

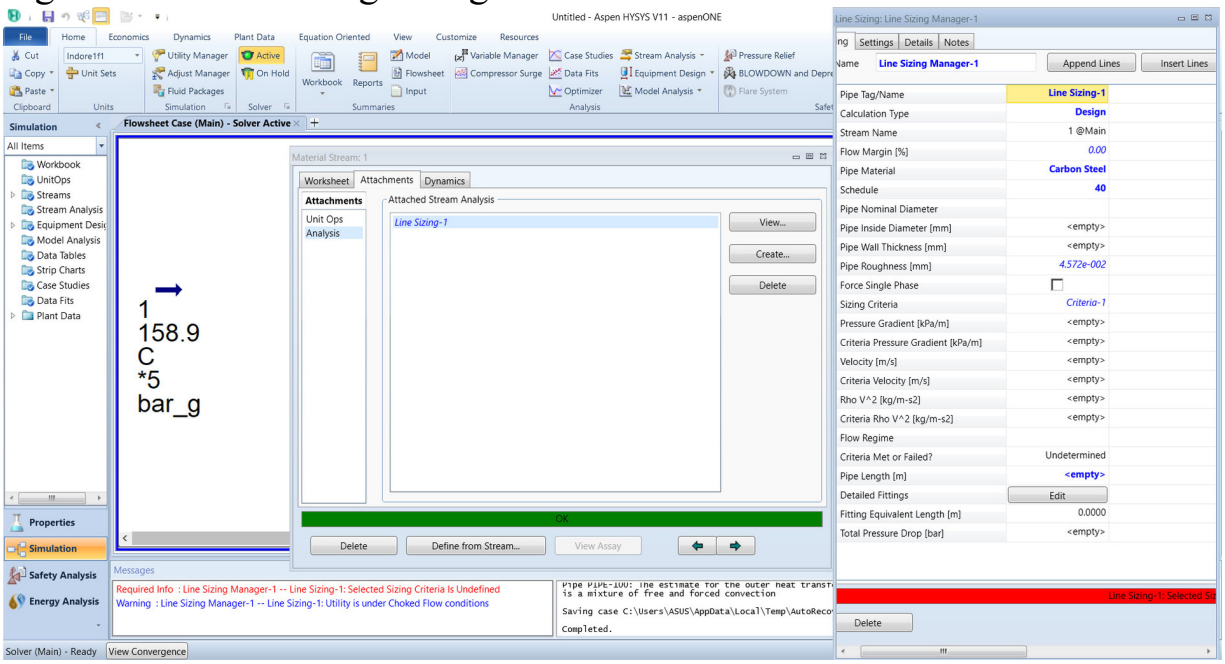
Click on “Create\Line Sizing Manager”.

Figure 5.31: Line sizing manager



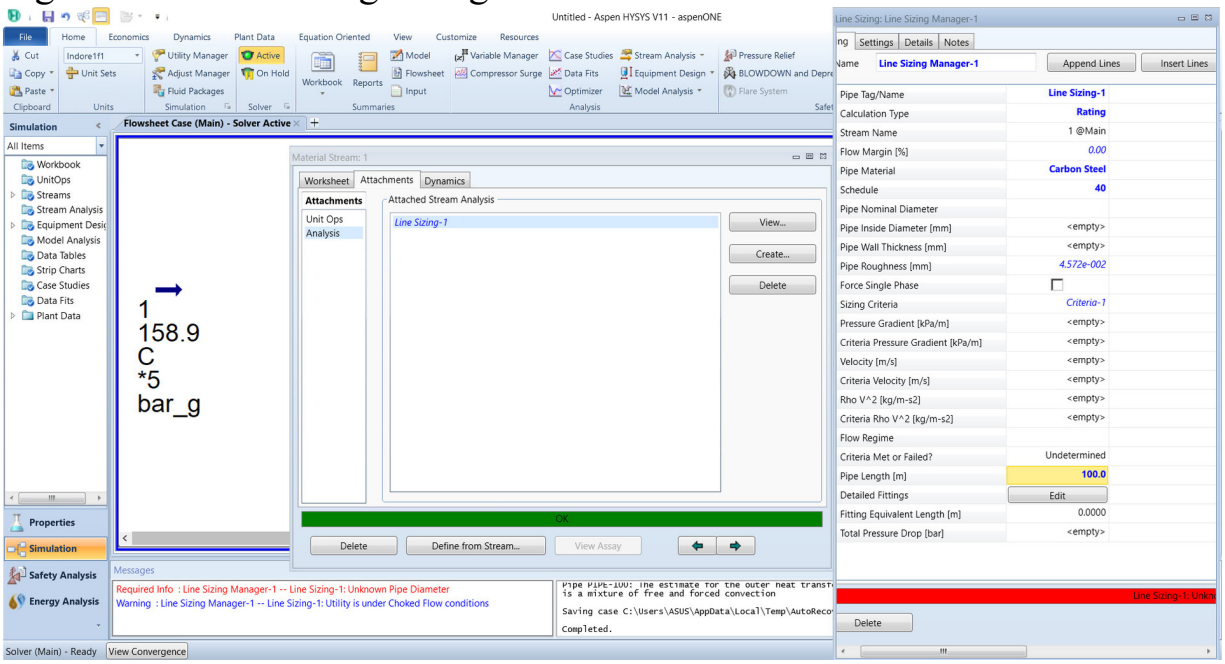
Below window will open.

Figure 5.32: Line sizing manager



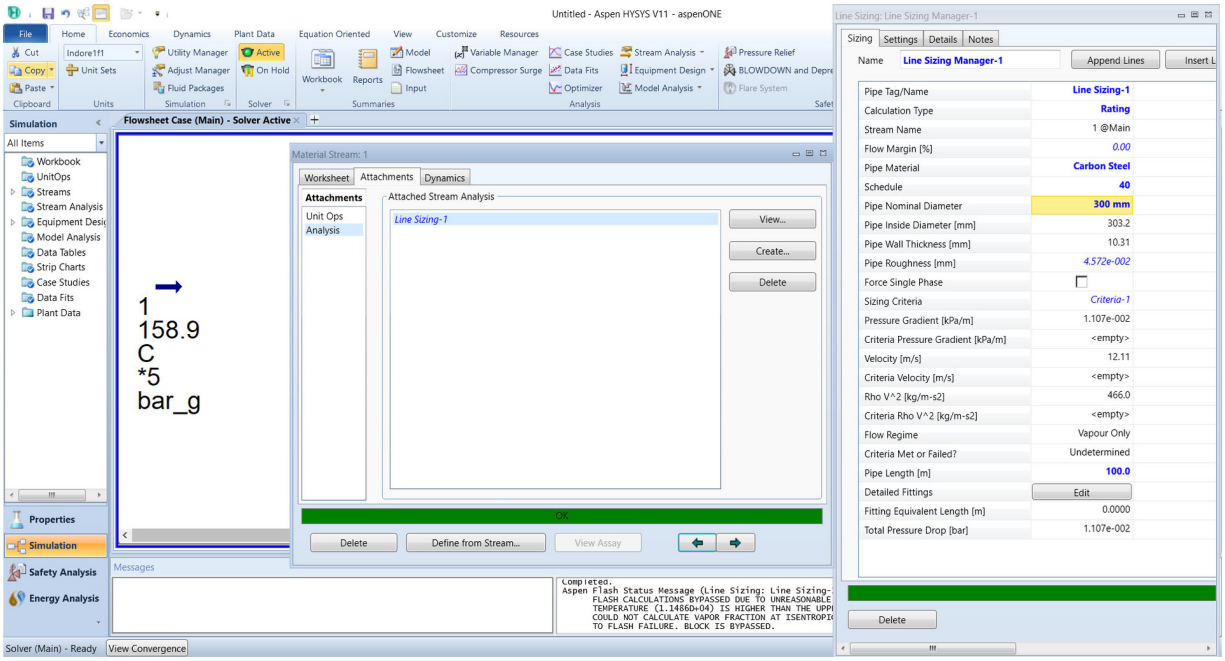
Select “Calculation Type” as “Rating”.

Figure 5.33: Line sizing manager



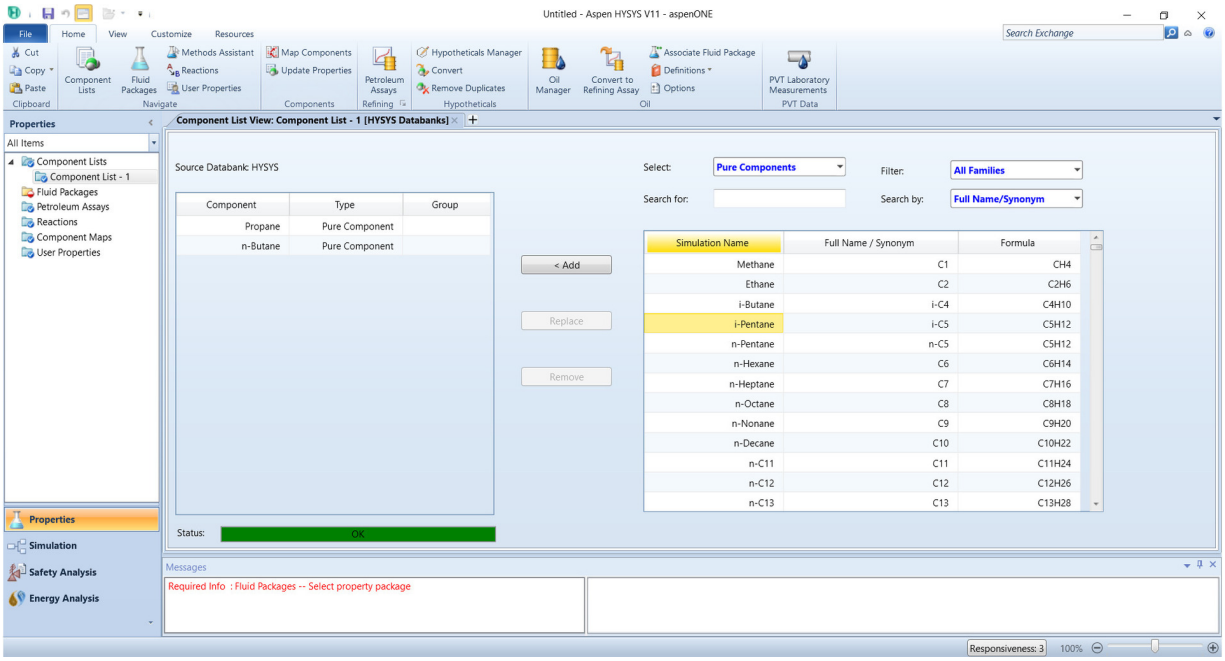
Choose “Pipe Nominal Diameter” as 300 mm and check if velocity is as per API 14E (< 18 m/s for vapor). Velocity calculated for 300 mm NB pipe is 12.1 m/s. Also, provide “Pipe Length” as 100 m (same as earlier problem).

Figure 5.34: Line sizing manager



Let’s now carry out two-phase line sizing. The material stream shall be 5% (mole) Propane and 95% (mole) n-Butane at 7 barg and 66 oC. The mass flow rate is 1000 kg/h.

Figure 5.35: Two phase line sizing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Fluid Packages
Methods Assistant
Reactions
User Properties
Navigate
Map Components
Update Properties
Components
Petroleum Assays
Refining
Hypotheticals Manager
Convert
Remove Duplicates
Hypotheticals
Oil Manager
Convert to Refining Assay
Associate Fluid Package
Definitions
Options
Oil
PVT Laboratory Measurements
PVT Data
Properties
All Items
Component Lists
Component List - 1
Fluid Packages
Petroleum Assays
Reactions
Component Maps
User Properties
Component List View: Component List - 1 [HYSYS Databanks]
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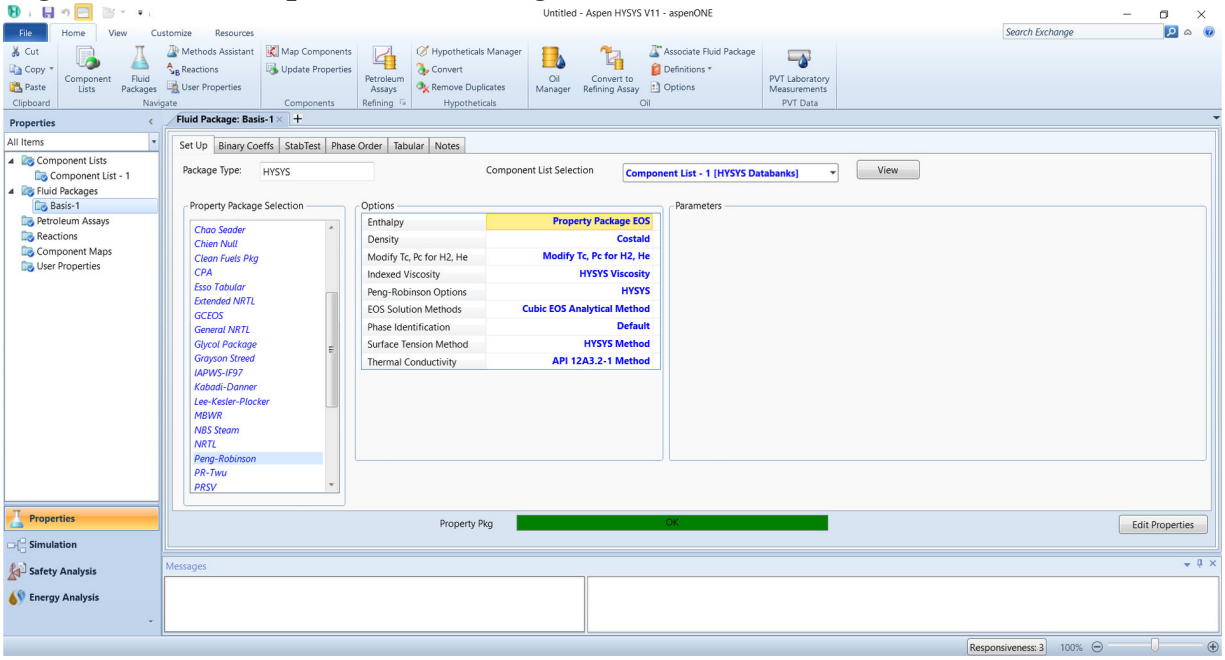
Component
Type
Group
Propane
Pure Component
n-Butane
Pure Component
< Add
Replace
Remove
Select:
Pure Components
Filter:
All Families
Search for:
Search by:
Full Name/Synonym
Simulation Name
Full Name / Synonym
Formula
Methane
C1
CH4
Ethane
C2
C2H6
i-Butane
i-C4
C4H10
i-Pentane
i-C5
C5H12
n-Pentane
n-C5
C5H12
n-Hexane
C6
C6H14
n-Heptane
C7
C7H16
n-Octane
C8
C8H18
n-Nonane
C9
C9H20
n-Decane
C10
C10H22
n-C11
C11
C11H24
n-C12
C12
C12H26
n-C13
C13
C13H28
Status:
OK
Properties
Simulation
Safety Analysis
Energy Analysis
Messages
Required Info : Fluid Packages -- Select property package
Responsiveness: 3
100%

The fluid package used is Peng Robinson.

Figure 5.36: Two phase line sizing



BIP are available for these components.

Figure 5.37: Two phase line sizing

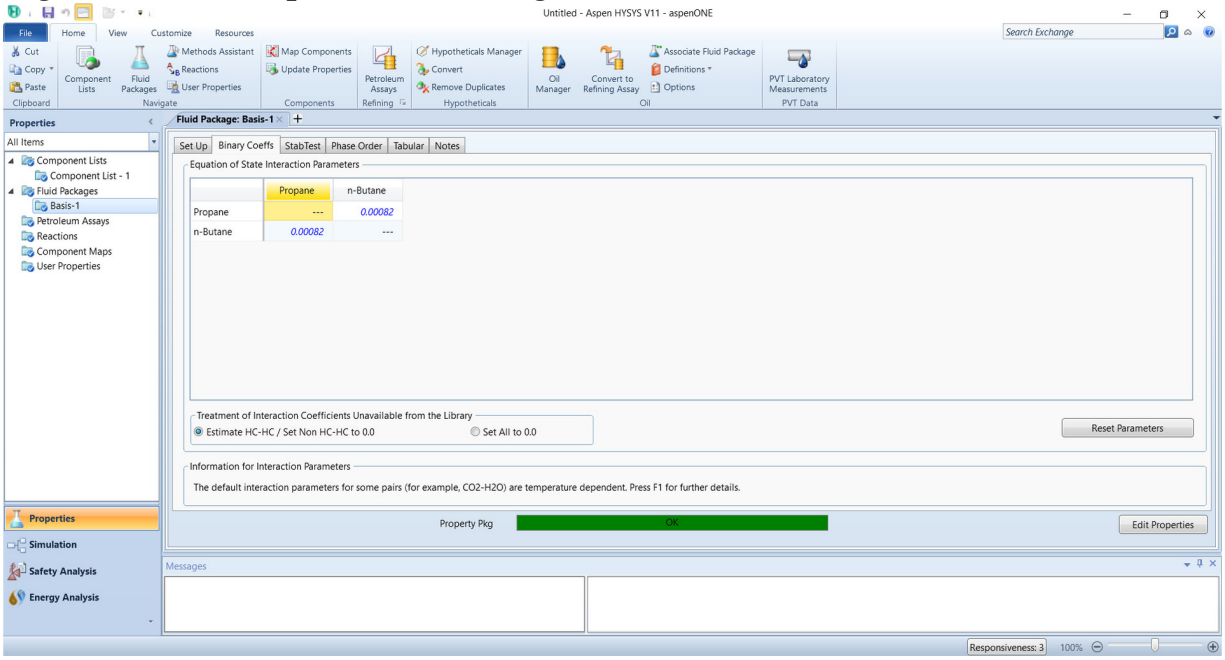


Figure 5.38: Two phase line sizing

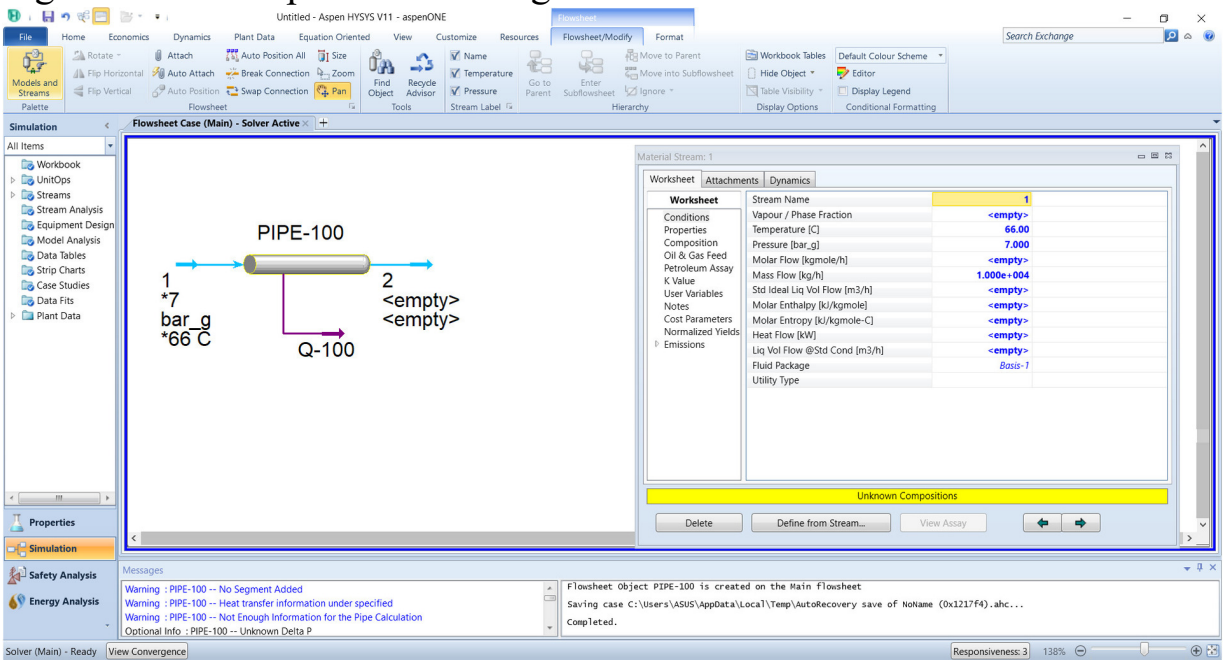
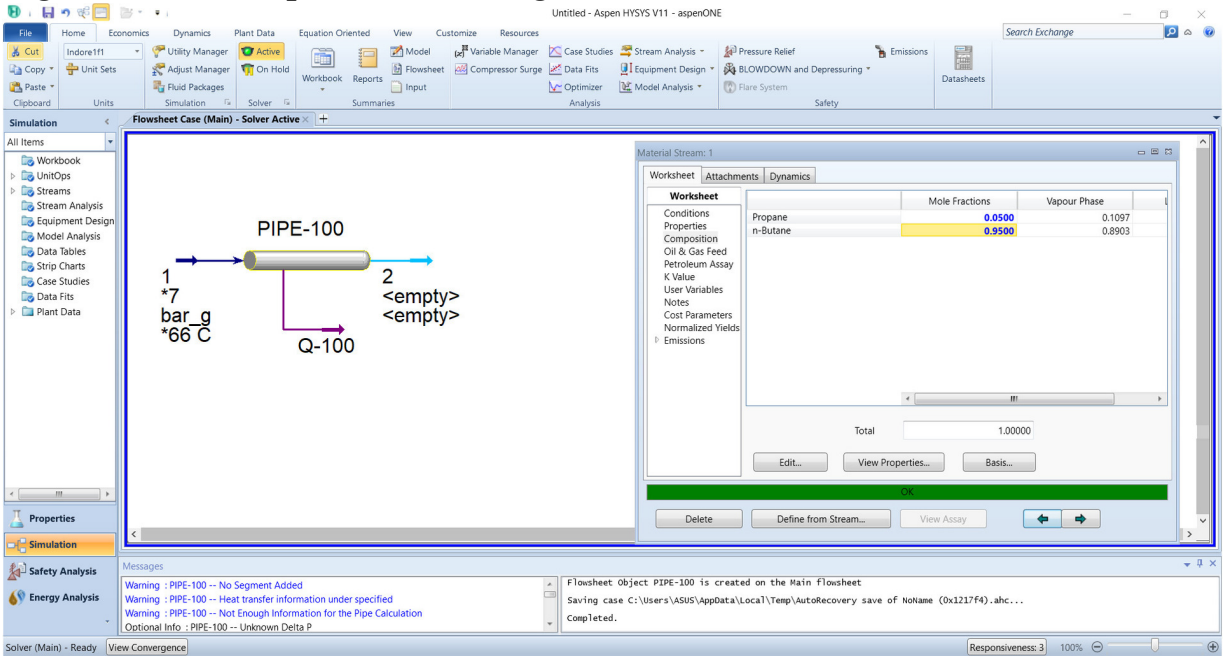
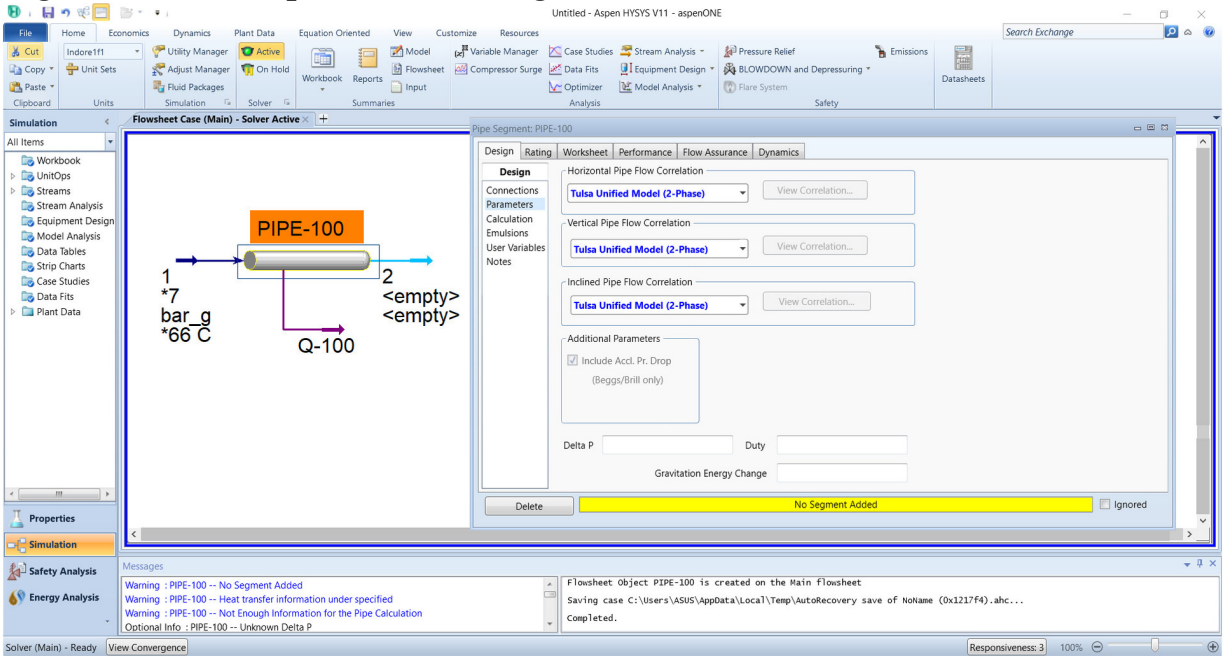


Figure 5.39: Two phase line sizing



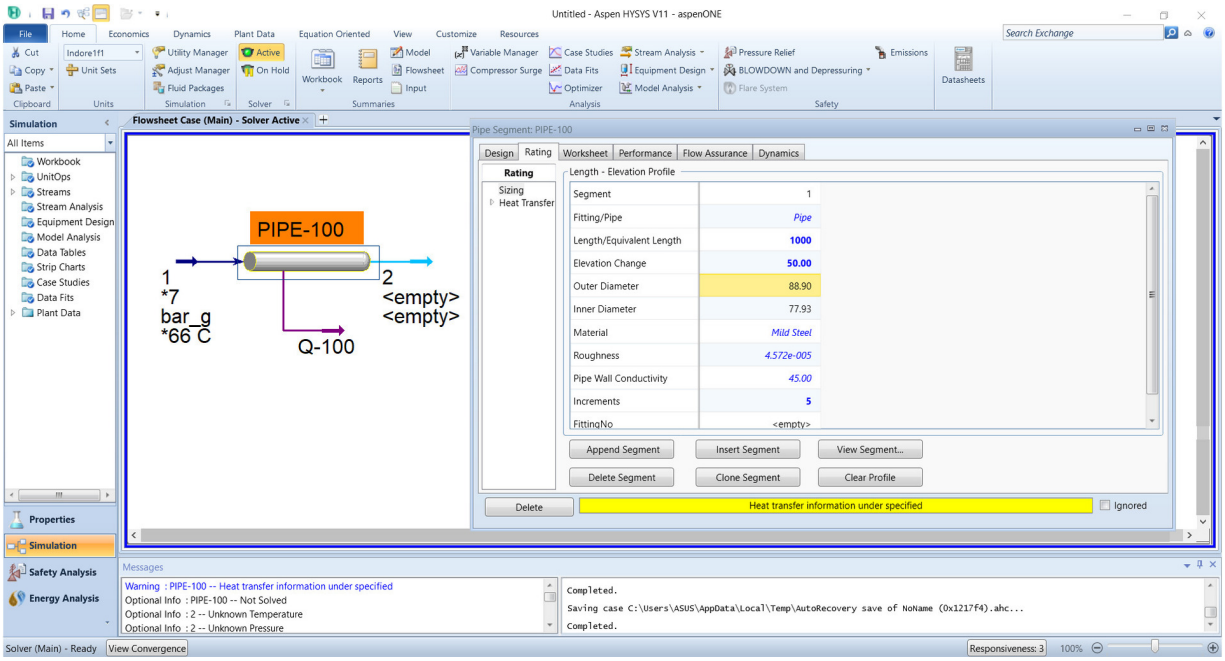
Double click on Pipe segment Select flow correlations as shown below.

Figure 5.40: Two phase line sizing



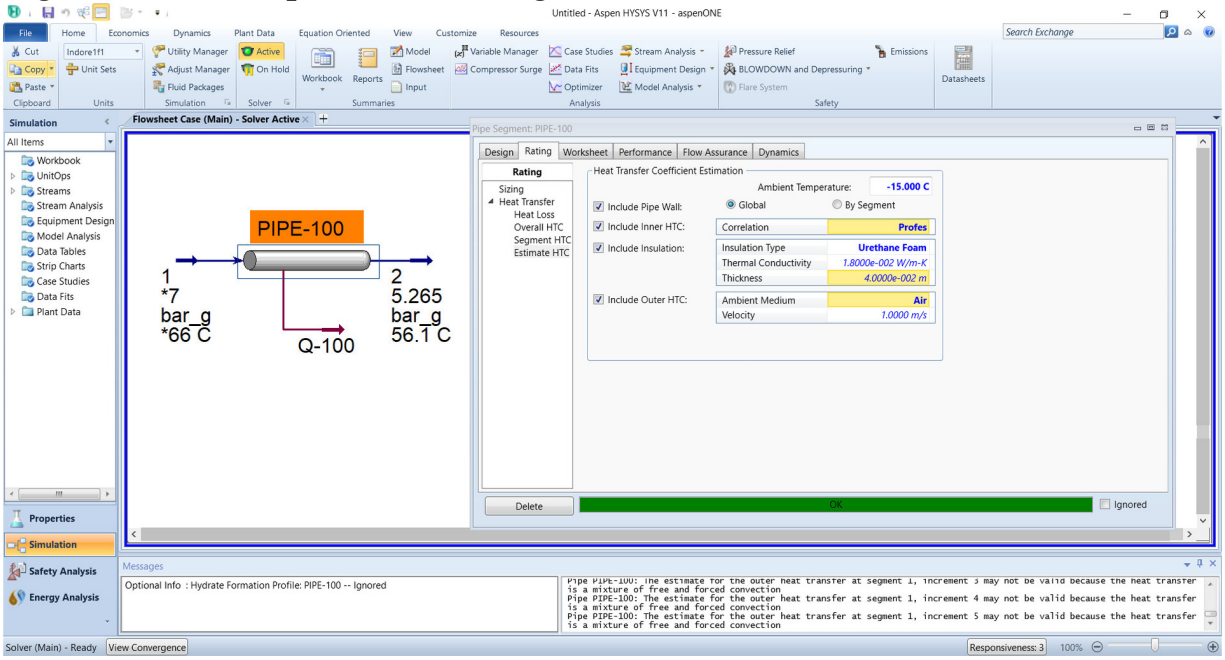
Pipe length is 1000 m with an elevation rise of 50 m. Pipe size selected is 80 mm NB.

Figure 5.41: Two phase line sizing



Heat transfer data is as below;

Figure 5.42: Two phase line sizing



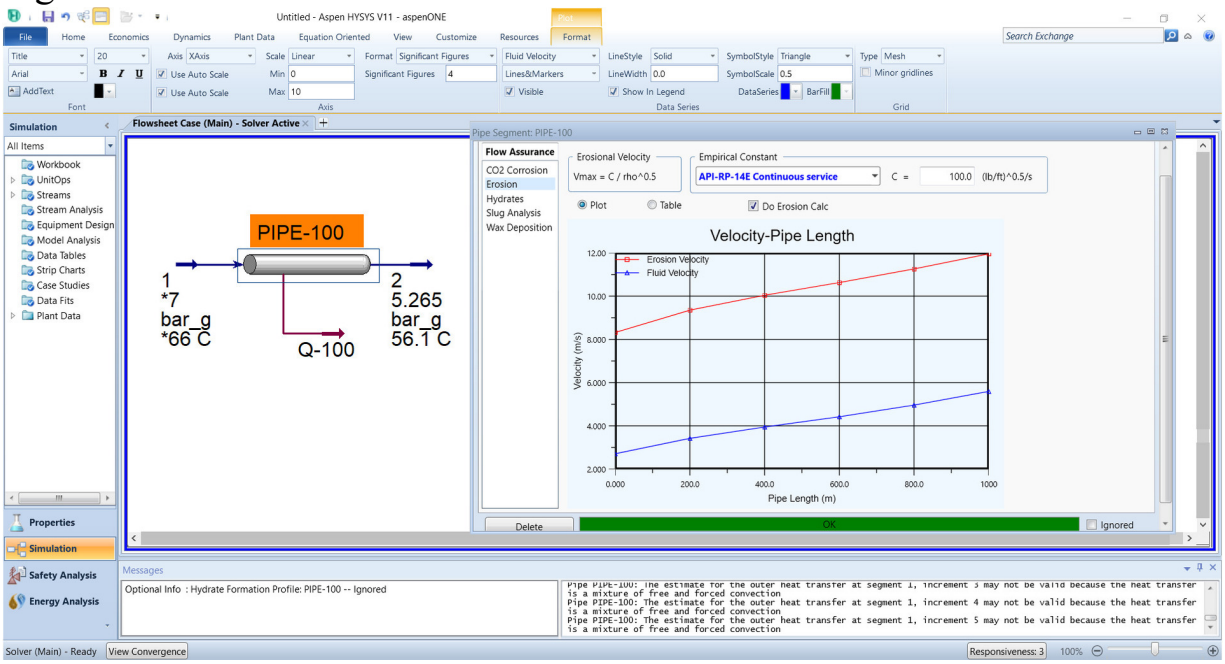
Select “Flow Assurance” tab. Then choose the “Erosion” tab. Check “Do Erosion Calc” box. It will calculate fluid velocity and erosion velocity as per API 14E. Fluid velocity should be comfortably less than erosion velocity. For pipe size of 80 mm NB, fluid velocity is less than erosion velocity for the given pipe length. Hence 80 mm NB is adequate for this application.

$$V_e = \frac{c}{\sqrt{\rho_m}}$$

Where,

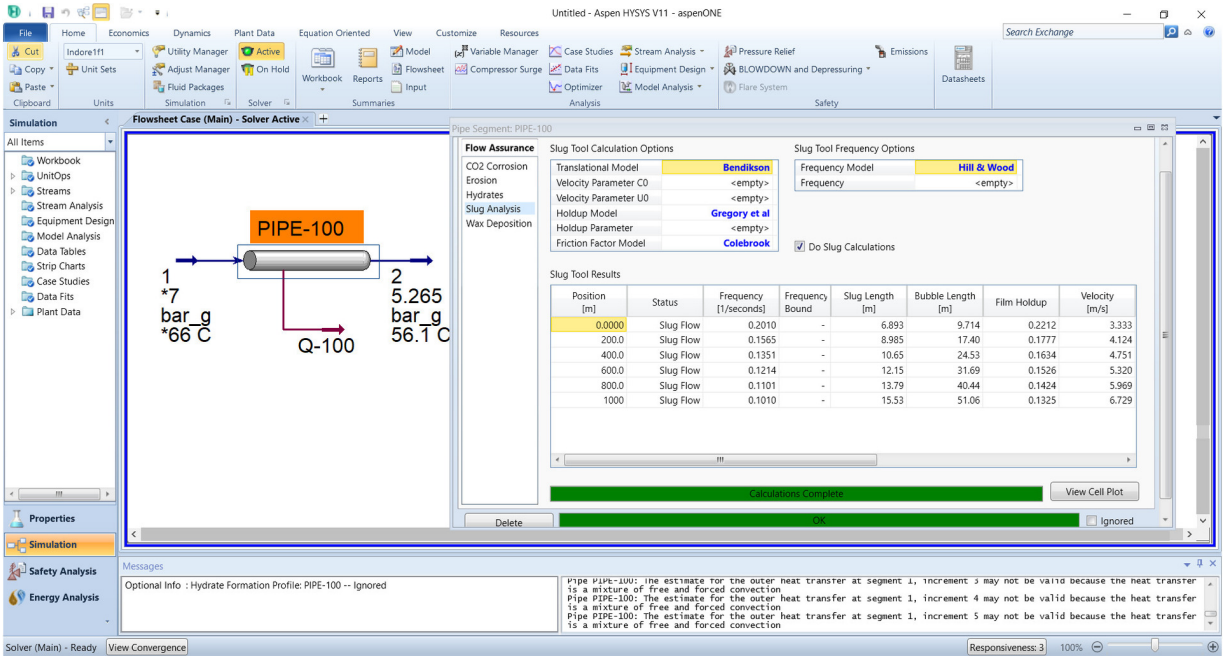
- V_e : Fluid erosional velocity, ft/s
- C : Empirical constant (c=100 for continuous system and c=125 for intermittent services)
- ρ_m : Gas/liquid mixture density at operating conditions

Figure 5.43: Flow assurance



We can also carry out Slug analysis in HYSYS. In this example for the first segment of 200 m, it can be noted that slug length is about 7 m followed by a bubble length of 10 m. This means that there is going to be a very serious slug problem in this pipe.

Figure 5.44: Flow assurance



In cases of intermittent flow, a higher design velocity might be considered as the cumulative effect of the higher velocities might not be an issue.

Exercises

a. Mr. Ravindra is designing a cooling water system. Can you find adequate line size for pump suction and discharge line, if the flow rate is 75 m^3 /h? What input parameters will be required to solve this problem? Can you make a reasonable assumption in this case?
b. Ms. Kanchan needs to decide the natural gas line size. Operating pressure, temperature, and flow

rate are 5 barg, 40 ° C, and 2500 kg/h respectively. Can you find out the line size?

c. Mr. Sundararajan is finding it difficult to select an adequate line size for 2 phase flow (methane and decane) at 3 barg pressure and 5% vapor content. Please help him.
d. Mr. Abdul is interested in finding out energy loss in INR in your city, if 5 barg saturated steam line (6", 500 m, 12 TPH) is kept uninsulated. What do you think his approach would be?
e. Mr. Nilesh is presenting his paper on flow-induced vibration. What this paper should contain?
f. Mr. Akshay wants to learn about surge analysis. Make a presentation file to help him learn.

CHAPTER 6: PUMP SIZING USING ASPEN HYSYS

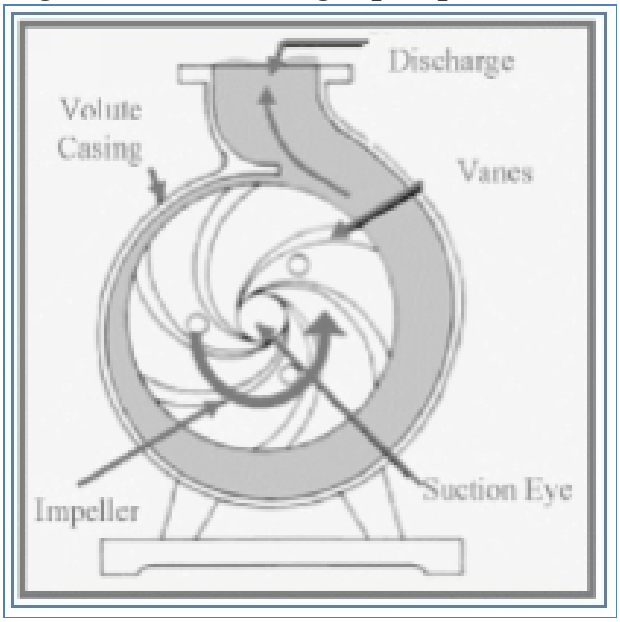
A pump is a mechanical device used to move fluids from one destination to another. Fluids can mean either liquids, gases, or a combination of liquids and solids (slurries).

Types of Pumps

Centrifugal Pump

A centrifugal pump is a roto-dynamic pump that uses a rotating impeller to increase the pressure and flow rate of a fluid. Centrifugal pumps are the most common type of pump used to move liquids through a piping system. The fluid enters the pump impeller along or near to the rotating axis and is accelerated by the impeller, flowing radially outward or axially into a diffuser or volute chamber, from where it exits into the downstream piping system. Centrifugal pumps are typically used for large discharge through smaller heads.

Figure 6.1: Centrifugal pump



Positive Displacement Pump

A positive displacement (PD) pump causes a fluid to move by trapping a fixed amount of it and then forcing (displacing) that trapped volume into the discharge pipe.

Some positive displacement pumps work using an expanding cavity on the suction side and a decreasing cavity on the discharge side. Liquid flows into the pump as the cavity on the suction side expands and the liquid flows out of

the discharge as the cavity collapses. The volume is constant given each cycle of operation.

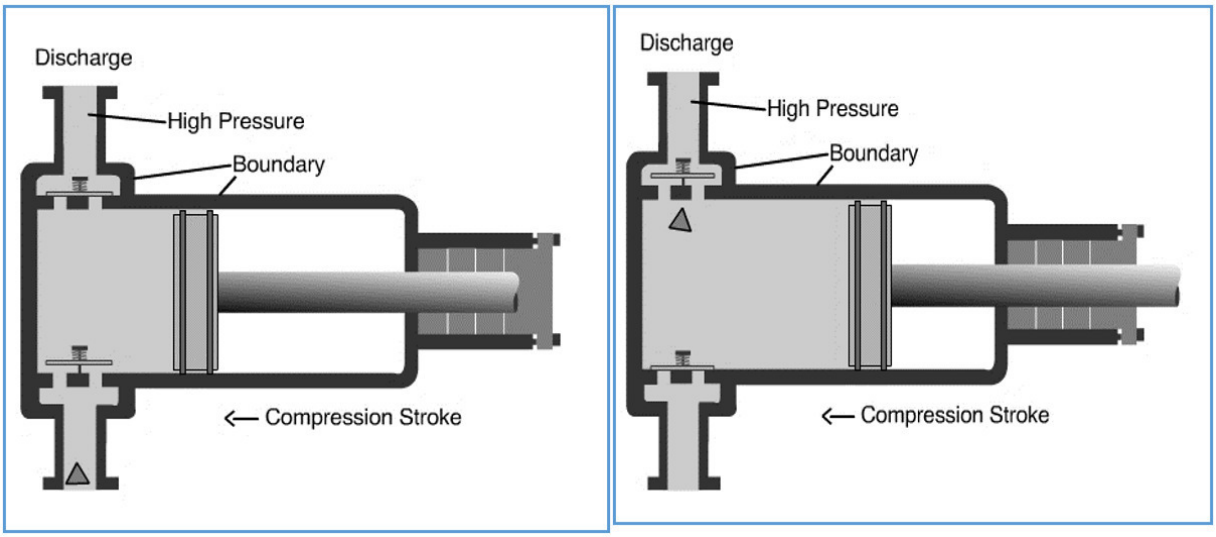
PD Pump behavior and safety

Positive displacement pumps, unlike centrifugal or roto-dynamic pumps, will in theory produce the same flow at a given speed (RPM) no matter what the discharge pressure. Thus, positive displacement pumps are "constant flow machines". However, due to a slight increase in internal leakage as the pressure increases, a truly constant flow rate cannot be achieved.

A positive displacement pump must not be operated against a closed valve on the discharge side of the pump, because it has no shut-off head like centrifugal pumps. A positive displacement pump operating against a closed discharge valve will continue to produce flow and the pressure in the discharge line will increase, until the line bursts or the pump is severely damaged, or both.

A relief or safety valve on the discharge side of the positive displacement pump is therefore necessary. The relief valve can be internal or external. The pump manufacturer normally has the option to supply internal relief or safety valves. The internal valve should in general only be used as a safety precaution, an external relief valve installed in the discharge line with a return line back to the suction line or supply tank is recommended.

Figure 6.2: PD pump



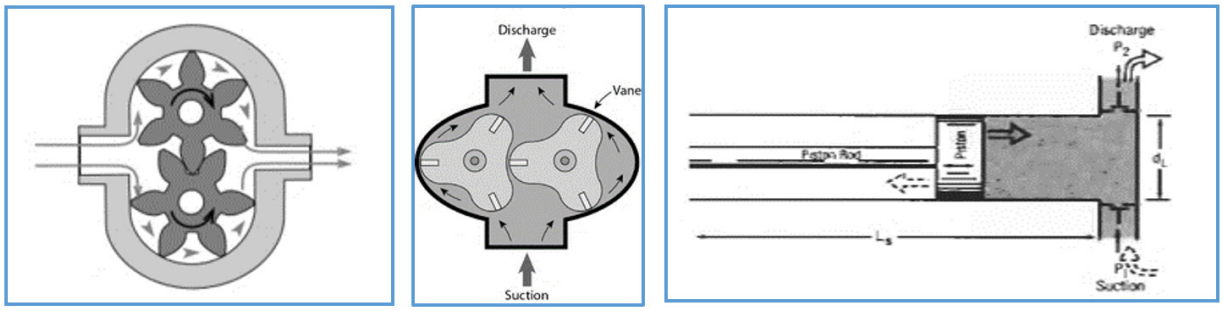
PD Pump Types

A positive displacement pump can be further classified according to the mechanism used to move the fluid:

- Rotary-type positive displacement: internal gear, screw, shuttle block, flexible vane or sliding vane, circumferential piston, or liquid ring vacuum pumps
- Reciprocating-type positive displacement: piston or diaphragm pumps.

Some illustrations for PD pumps are provided below:

Figure 6.3: PD pump



Design Flow Rate

The pump design flow rate shall include a minimum 10% margin on the maximum flow rate except in the following cases:

- Intermittent service
- Minimum flow bypass using orifice plate, in which case pump design flow rate shall be the process flow rate (as per heat and material balance) plus the minimum flow bypass rate.
- For reflux pumps, the design flow rate shall be at least 120% of the operating flow rate.

Minimum Flow By-Pass

API STD 610 defines the following two minimum flows for centrifugal pumps:

- Minimum continuous stable flow:

It is the lowest flow at which the pump can operate without exceeding acceptable vibration limits.

- Minimum continuous thermal flow:

It is the lowest flow at which the pump can operate without its operation being impaired by the temperature rise of the pumped liquid.

Pump operations below these points can cause shaft vibration and reduce the mechanical seal life.
To avoid such problems, high capacity pumps shall be provided with a minimum flow bypass with the flow controlled by a restriction orifice or a control valve. If a restriction orifice is used, the operating flow rate of the pump should be increased to account for the continuous bypass flow.

The sizing of the minimum bypass flow circuit shall be based on the higher of the two minimum flows.
If it is not possible to provide a minimum flow circulation line, then adequate instrument protection shall be provided to prevent pump operation below the minimum allowable flow rate.

Centrifugal Pumps Guidelines

- The head/capacity characteristic curve shall continuously rise as flow is reduced to shut-off (or zero flow). The minimum recommended rise is 12% concerning the rated point. For parallel

operation, the head rise shall be between a minimum of 15% to a maximum of 20%.

- The pump should be capable of a head increase at the design conditions by installing a larger impeller. The rated impeller diameter shall be between 85% and 90% of the maximum impeller diameter as per vendor datasheet.
- Pumps should not be operated below the minimum safe continuous flow as provided by the pump manufacturer.
- The installed power shall be adequate for maximum impeller diameter plus 120% of rated capacity.

Positive Displacement Guidelines

- Reciprocating/Controlled volume pumps are capable of handling very low flows (< 0.227 m^3 /h) and are typically used for chemical injection duties in oil and gas installations.
- Reciprocating pumps should comply with API STD 674.
- Metering (or controlled volume) pumps should comply with API STD 675.
- Some selection guidelines for controlled volume pumps are as follows:
    - "Packed Plunger Pumps" are generally recommended for pressures above 340 barg
    - Diaphragm pumps are generally recommended for pressures up to 340 barg and temperatures up to 150 °C.
    - Double Diaphragm Pumps of the sandwich-type construction are generally recommended.

    - Double Diaphragm pumps should be used when:
- Dilution of the product flow by hydraulic oil is not acceptable.
- Pumped fluid is extremely hazardous and/or toxic.
- Process leakage to the atmosphere is unacceptable
- Process fluid must not come in contact with air

Capacity Adjustment for Metering Pumps

- Pump capacity shall be adjusted by changing the actual or effective stroke length or the pump stroking speed.
- The pump shall be capable of accepting manual or automatic capacity stroke control, either factory mounted or by field conversion.
- Integral pump devices used to vary capacity either manually or automatically shall be provided with a visual indication of capacity setting, shown as a percentage of the nameplate rated flow. Manual control shall include a locking device to positively retain the capacity setting.
- The direction of movement to increase or decrease pump flow shall be marked. All adjustment means and indicators shall be easily accessible with the pump installed.

Pressure Relief for Positive Displacement Pumps

Pressure relief will be provided at the pump discharge upstream of any isolating valve. Where the nature of the fluid would make relief valves unreliable, rupture discs should be used either alone or in conjunction with relief valves.

Pulsation Devices for Positive Displacement Pumps

When specified, the vendor will furnish pulsation suppression devices to be located at the pump suction and/or discharge connections. The following are three basic types of pulsation suppression devices:

- Volume bottles without internals.
- Pulsation dampeners using a gas-filled chamber isolated from the pumped fluid by a piston or elastomeric diaphragm.

- Pulsation filters and attenuators, including proprietary commercial designs based on acoustical suppression techniques.

Reference is made to API STD 675 for the above.

Further Explanation of Key Terms Related to Pumps

Net positive suction head available

Net positive suction head available (NPSHa) is the difference between the liquid pressure at the suction point of the pump and vapour pressure of the liquid. It is calculated by process engineer and is a aprt of information in pump datasheet that is provided to the pump vendor. Vendor selects the pump model based on NPSHa (given by the user) and NPSHr (a function of pump model). NPSHa should be greater than NPSHr by 1 m or more, as the best engineering practice.

The formula for calculating NPSHa is:

$NPSHa = P_{ps} - H_{vp} - H_a$

Where:

- P_{ps} = absolute pressure at the pump suction nozzle, meter of liquid
- H_{vp} = vapor pressure of the liquid at the pumping temperature, expressed in meter of liquid
- H_a = Acceleration head. Acceleration head needs to be considered only for reciprocating pumps. For centrifugal pumps and rotary pumps, the acceleration head is zero.

Net positive suction head required

Net positive suction head required (NPSHr) is the minimum pressure at the pump suction to keep the pump from cavitating which causes erosion, noise and pump damage. NPSHr is an intrinsic property of the pump and is determined by pump vendor experimentally.

Differential Pressure or Differential head

It is the difference between the discharge pressure and the suction pressure.

Pump Motor Brake Kilowatt

Pump requires a driver such as a motor or turbine to drive the pump. The Brake Horse Power (BHP) or Brake Kilowatt (BkW) needs to be provided by the process engineer for the selection of the driver by the electrical engineer. Expressed in equation form the BHP or BkW would be:

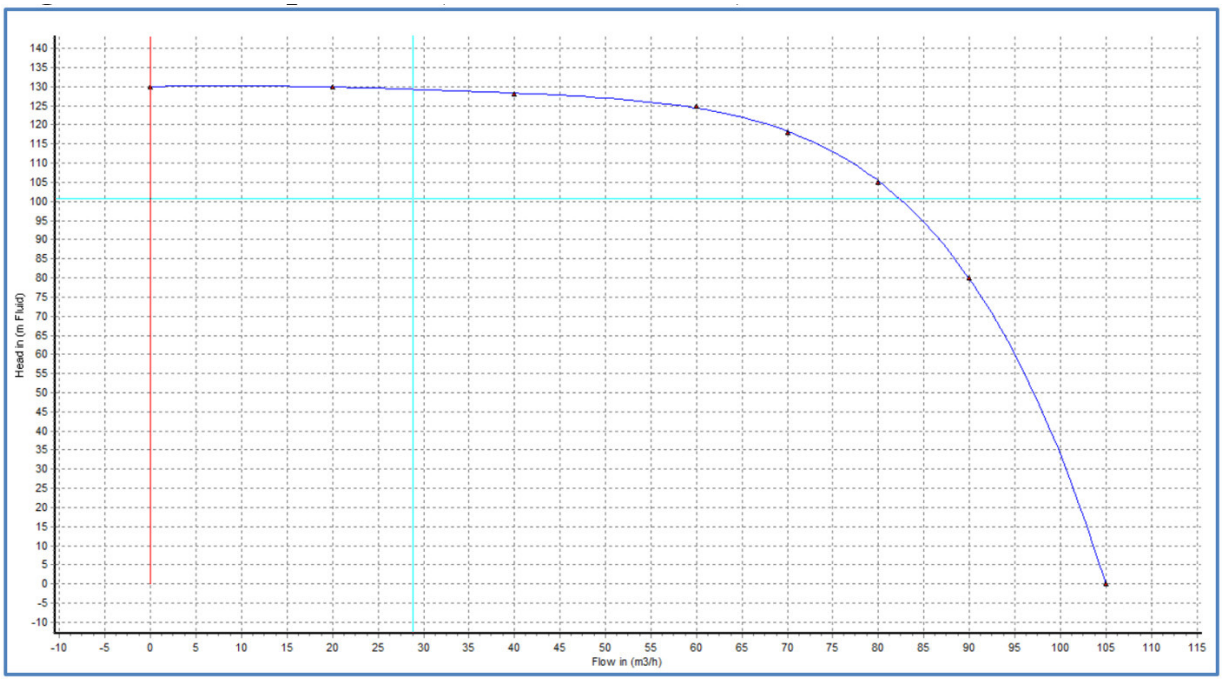
$$BkW = Q \times P_{d\,(rated)} \times \rho \times g / (3600 \times \eta)$$

Where:

- BkW : Brake Power in kW
- Q : Rated volume flow rate of the pump, m^3 /h
- P_d : Differential pressure, m
- P : Density of the liquid at the pumping temperature, kg/m^3
- g : Acceleration due to gravity, 9.81 m/s^2
- η : Pumping efficiency, expressed as a fraction

Typical Pump Curve

Figure 6.4: Pump curve (head vs flowrate)


Head in (m Fluid)
Flow in (m3/h)
140
135
130
125
120
115
110
105
100
95
90
85
80
75
70
65
60
55
50
45
40
35
30
25
20
15
10
5
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-5
-10
-10
-5
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15
20
25
30
35
40
45
50
55
60
65
70
75
80
85
90
95
100
105
110
115

Figure 6.5: Pump curve (head vs pump efficiency)

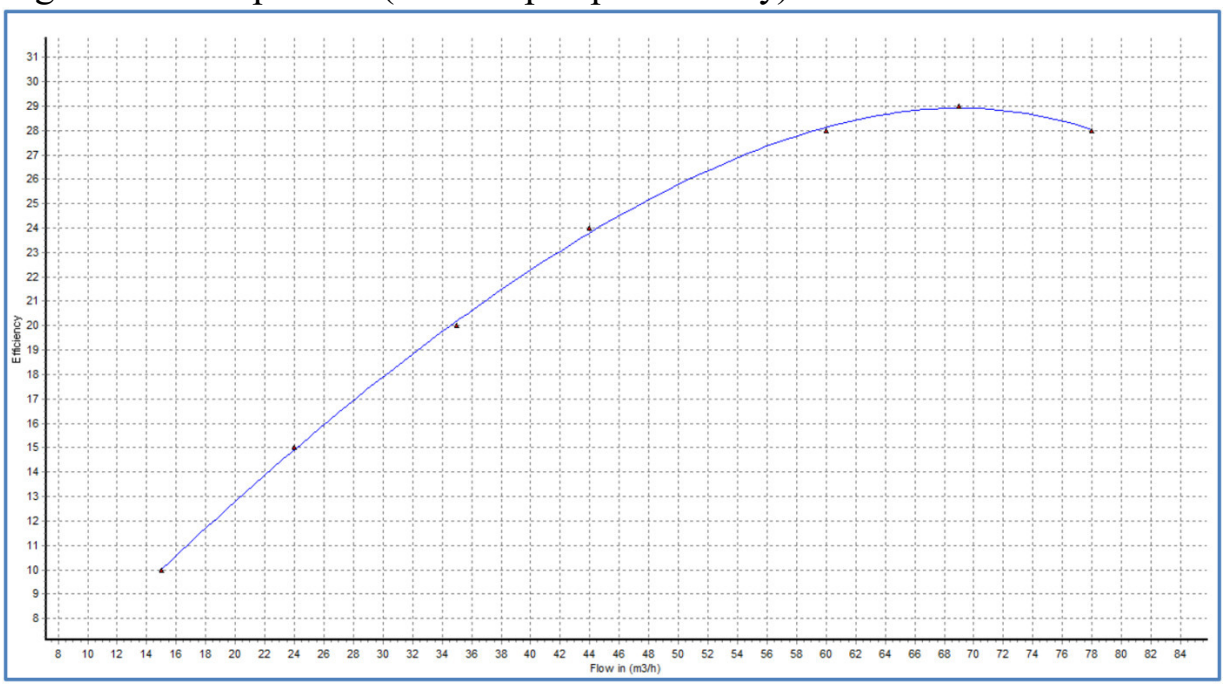
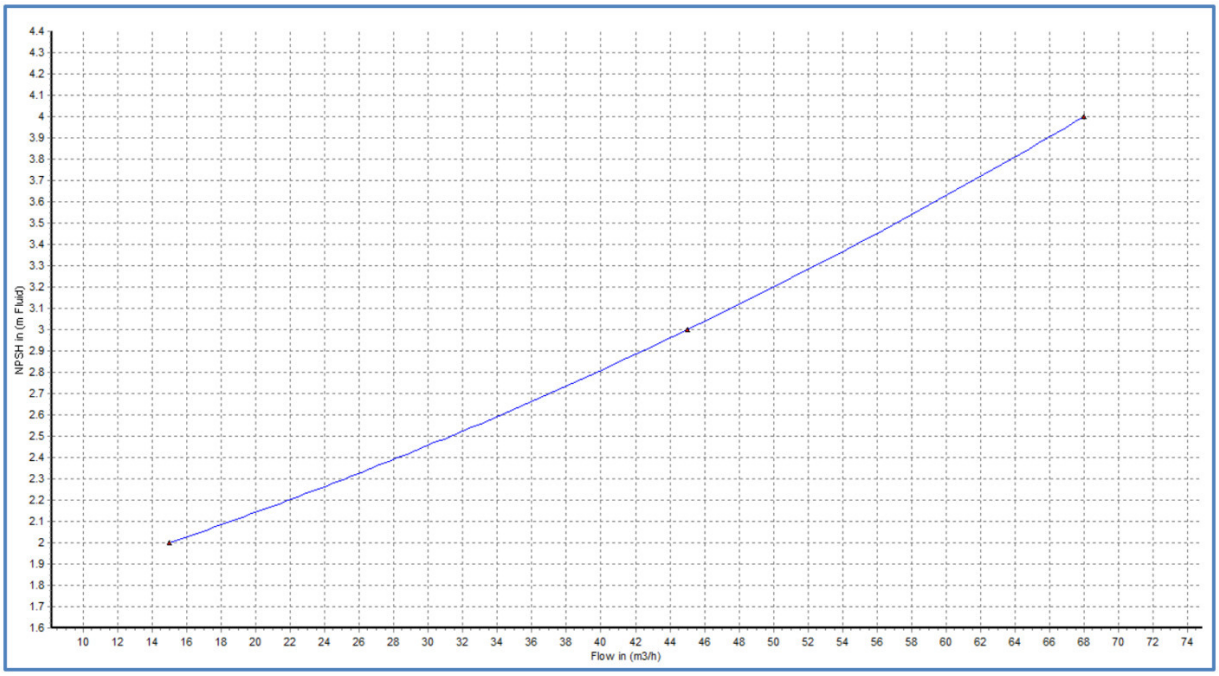


Figure 6.6: Pump curve (head vs NPSHr)



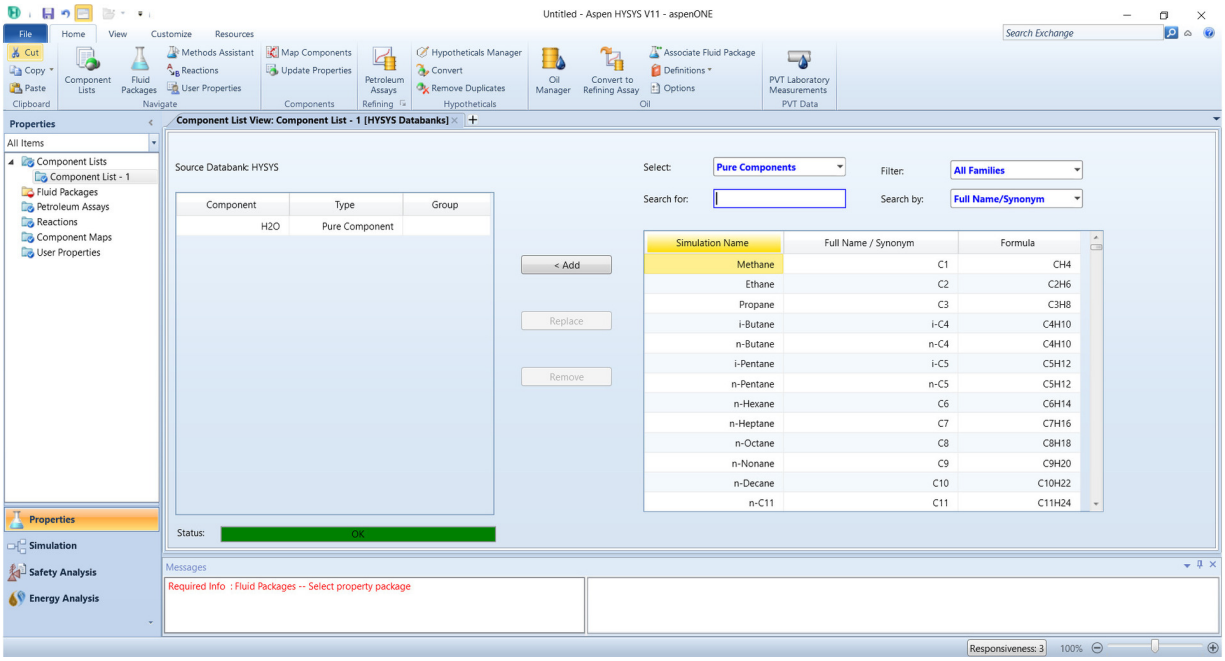
Simulation

Let's now solve a pump sizing problem using Aspen HYSYS.

Water at 30 ° C and 0 barg pressure is pumped to 3 barg pressure. The water flow rate is 30 m^3 /h.

To solve this problem, we need to first select water as a component and IAPWS as Fluid Package.

Figure 6.7: Component list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Fluid Packages
Methods Assistant
Reactions
User Properties
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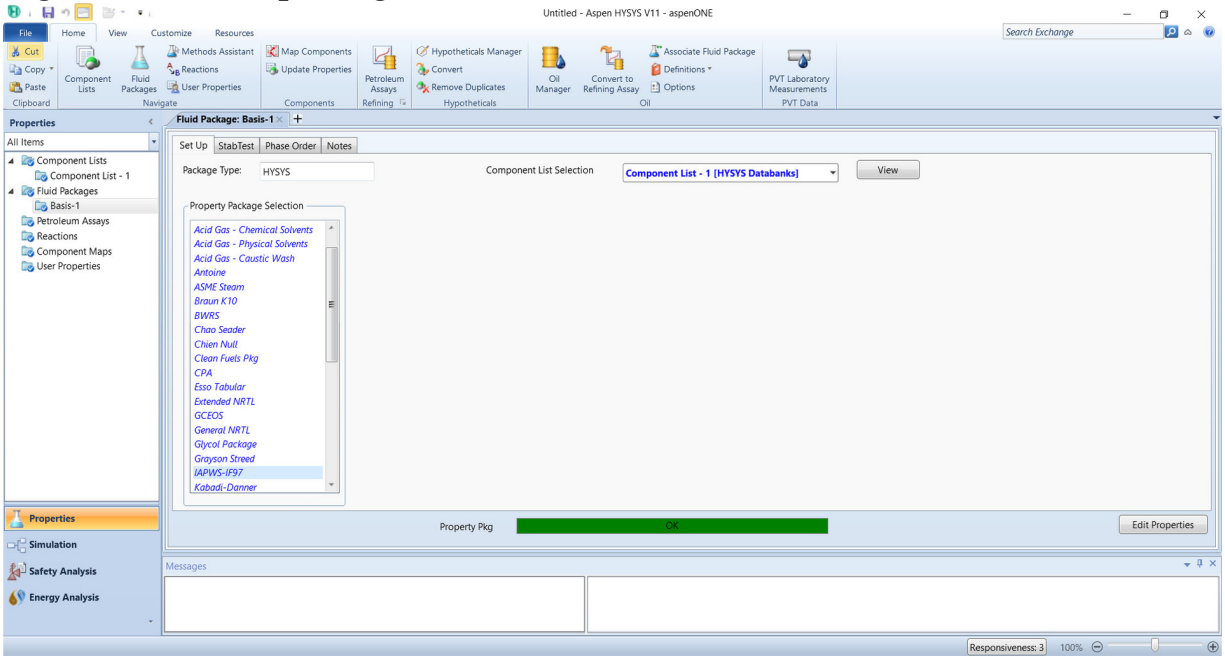
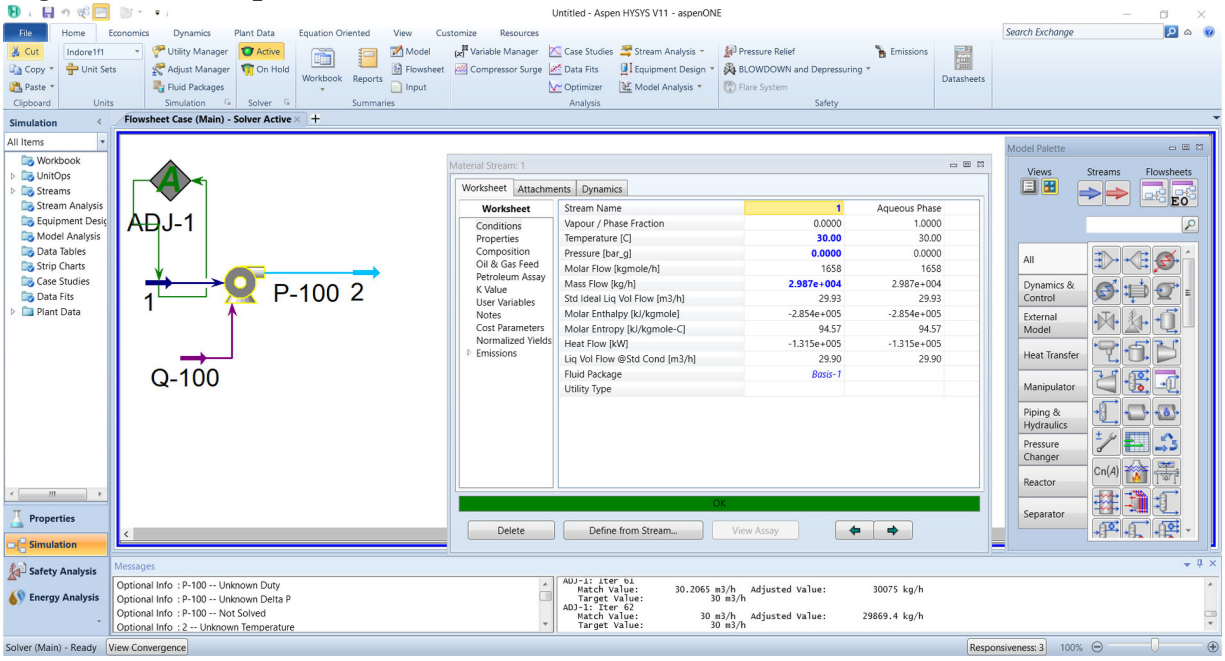
Map Components
Update Properties
Components
Petroleum Assays
Refining
Hypotheticals Manager
Convert
Remove Duplicates
Hypotheticals
Oil Manager
Convert to Refining Assay
Associate Fluid Package
Definitions
Options
Oil
PVT Laboratory Measurements
PVT Data
Properties
All Items
Component Lists
Component List - 1
Fluid Packages
Petroleum Assays
Reactions
Component Maps
User Properties
Component List View: Component List - 1 [HYSYS Databanks]
Source Databank: HYSYS
Component
Type
Group
H2O
Pure Component
Select:
Pure Components
Filter:
All Families
Search for:
Search by:
Full Name/Synonym
< Add
Replace
Remove
Simulation Name
Full Name / Synonym
Formula
Methane C1 CH4
Ethane C2 C2H6
Propane C3 C3H8
i-Butane i-C4 C4H10
n-Butane n-C4 C4H10
i-Pentane i-C5 C5H12
n-Pentane n-C5 C5H12
n-Hexane C6 C6H14
n-Heptane C7 C7H16
n-Octane C8 C8H18
n-Nonane C9 C9H20
n-Decane C10 C10H22
n-C11 C11 C11H24
Status:
OK
Properties
Simulation
Safety Analysis
Energy Analysis
Messages
Required Info : Fluid Packages -- Select property package
Responsiveness: 3
100%

Figure 6.8: Fluid package



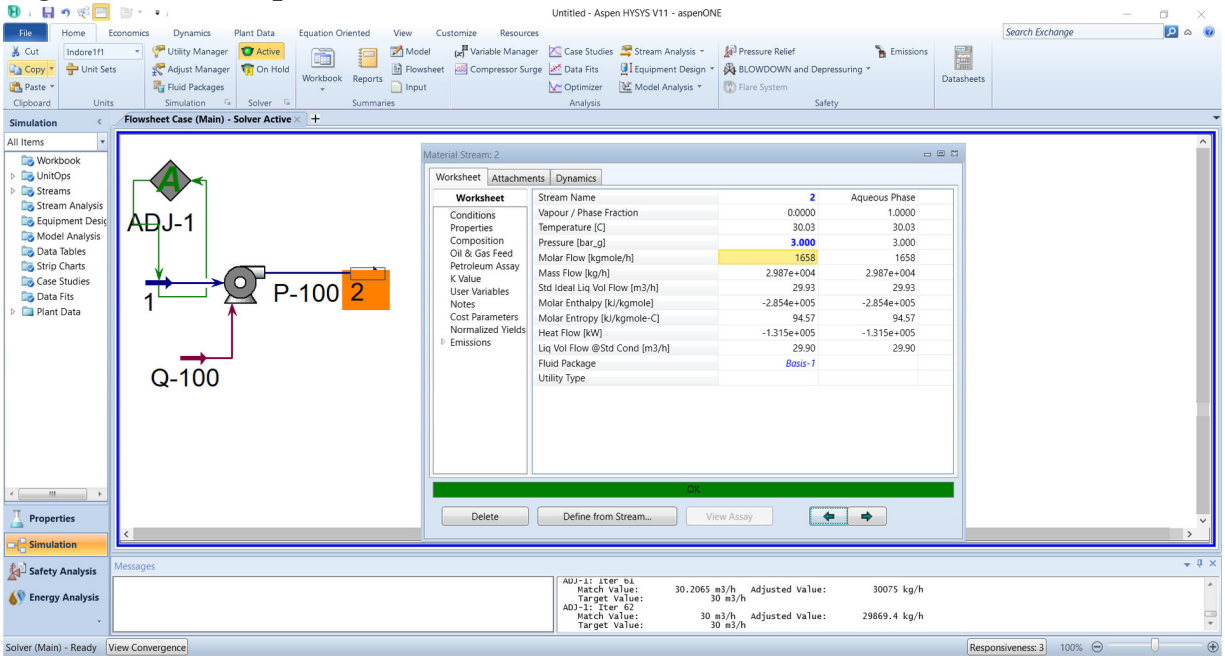
Use model palette to get 2 material streams and a pump, connect those items, and provide given data in the inlet stream as shown below. Use Adjust function to achieve 30 m^3 /h flow by varying the mass flow rate of stream "1".

Figure 6.9: Pump



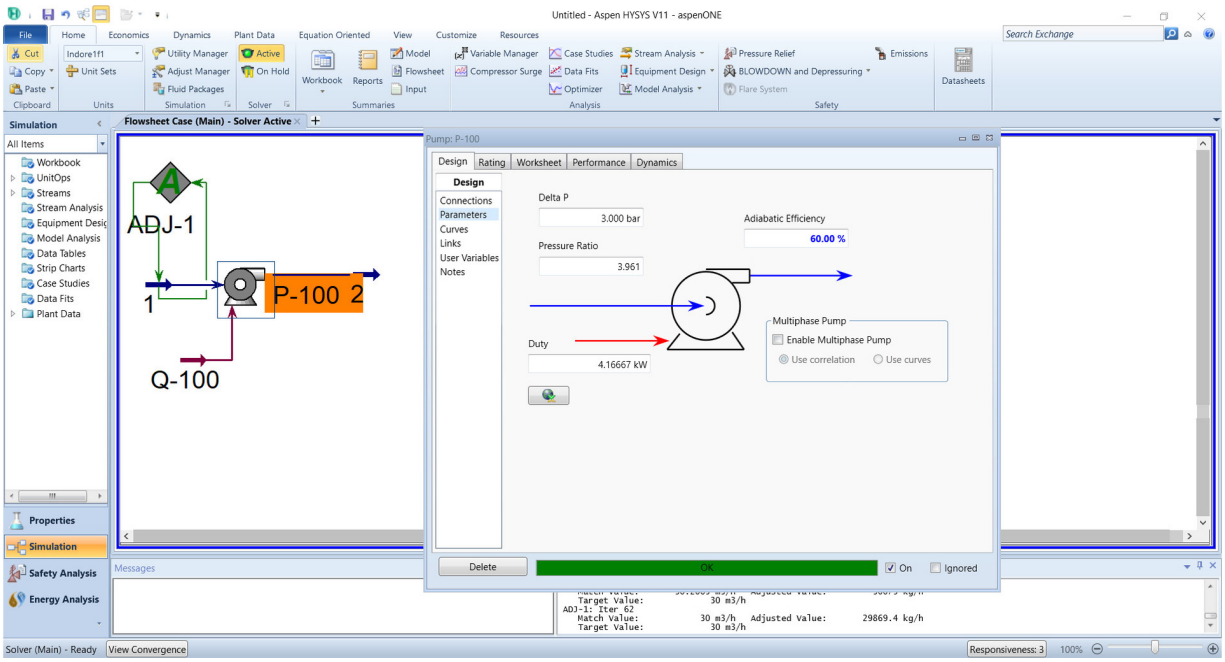
Provide outlet pressure as 3 barg.

Figure 6.10: Pump



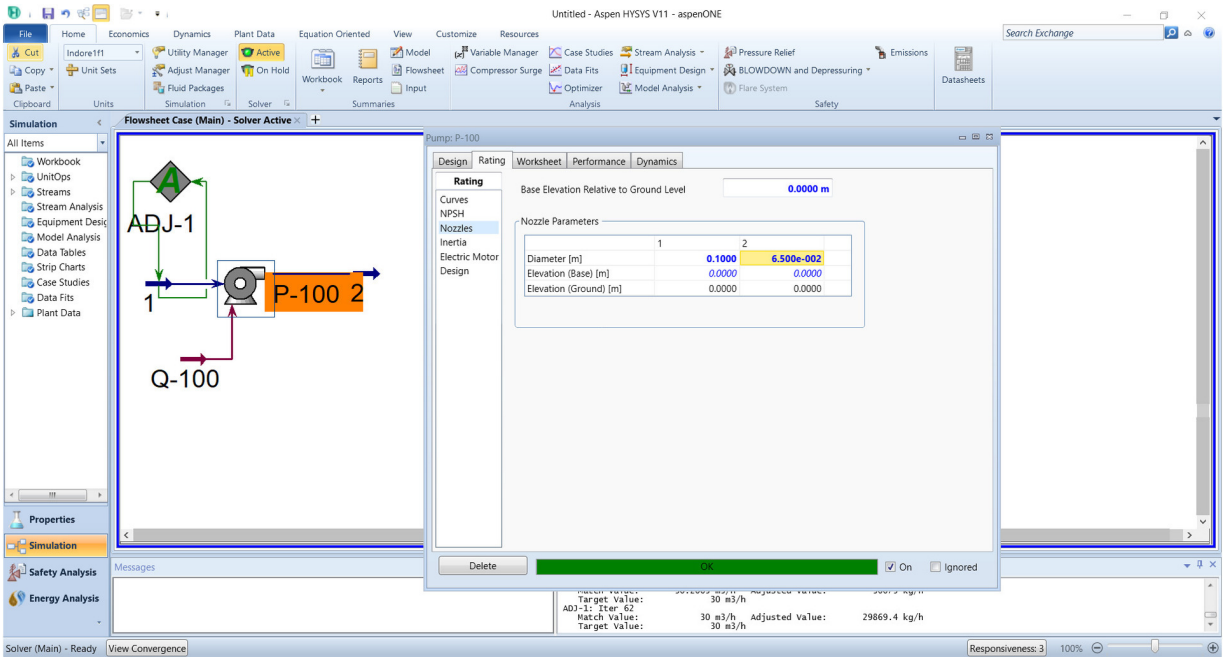
Double click on Pump P-100. In the Design\Parameter tab, default pump efficiency is given as 75%. Change it to 60%.

Figure 6.11: Pump



Select “Rating\Nozzles” Tab and change Inlet and Outlet pipe diameter as 0.1 m and 0.065 m (considering velocity criteria).

Figure 6.12: Pump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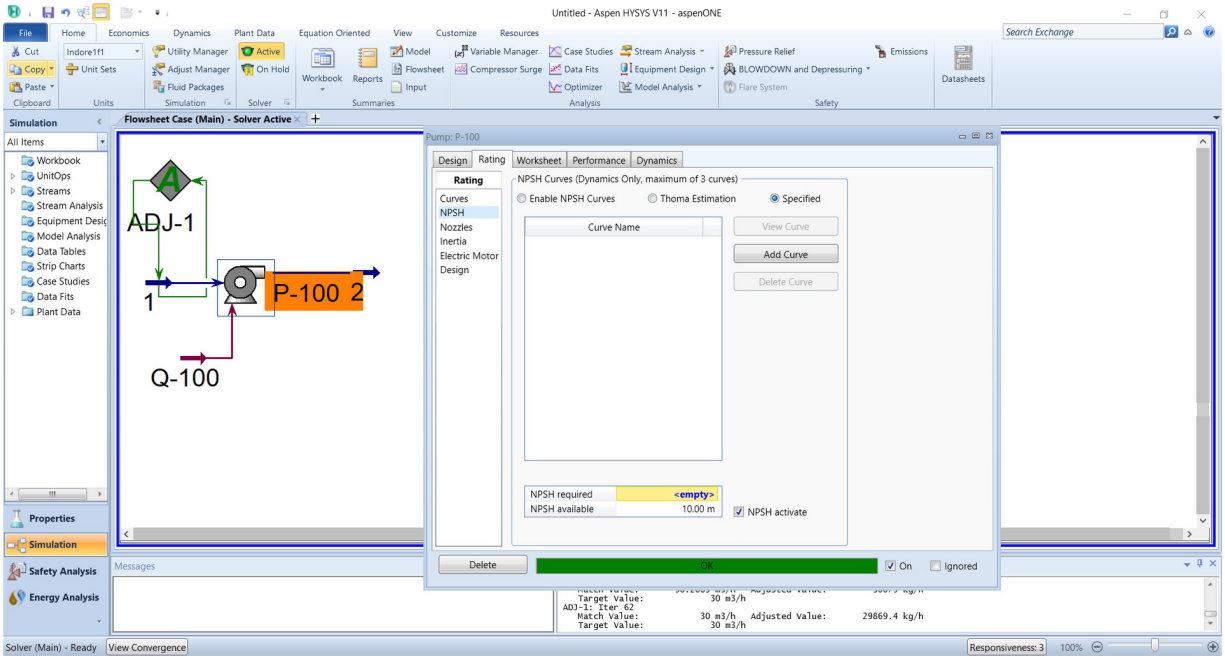
Units
Utility Manager
Adjust Manager
Fluid Packages
Simulation
Active
On Hold
Solver
Workbook
Reports
Model
Flowsheet
Input
Summaries
Variable Manager
Compressor Surge
Case Studies
Data Fits
Optimizer
Stream Analysis
Equipment Design
Model Analysis
Analysis
Pressure Relief
BLOWDOWN and Depressuring
Flare System
Emissions
Safety
Datasheets
Simulation
All Items
Workbook
UnitOps
Streams
Stream Analysis
Equipment Desig
Model Analysis
Data Tables
Strip Charts
Case Studies
Data Fits
Plant Data
Properties
Simulation
Safety Analysis
Energy Analysis
Flowsheet Case (Main) - Solver Active
ADJ-1
1
P-100
2
Q-100
Pump: P-100
Design
Rating
Worksheet
Performance
Dynamics
Rating
Curves
NPSH
Nozzles
Inertia
Electric Motor
Design
Base Elevation Relative to Ground Level
0.0000 m
Nozzle Parameters
1
2
Diameter [m]
0.1000
6.500e-002
Elevation (Base) [m]
0.0000
0.0000
Elevation (Ground) [m]
0.0000
0.0000
Delete
OK
On
Ignored
Messages
Target Value: 30 m3/h
ADJ-1: Iter 62
Match Value: 30 m3/h
Adjusted Value: 29869.4 kg/h
Target Value: 30 m3/h
Solver (Main) - Ready
View Convergence
Responsiveness: 3
100%

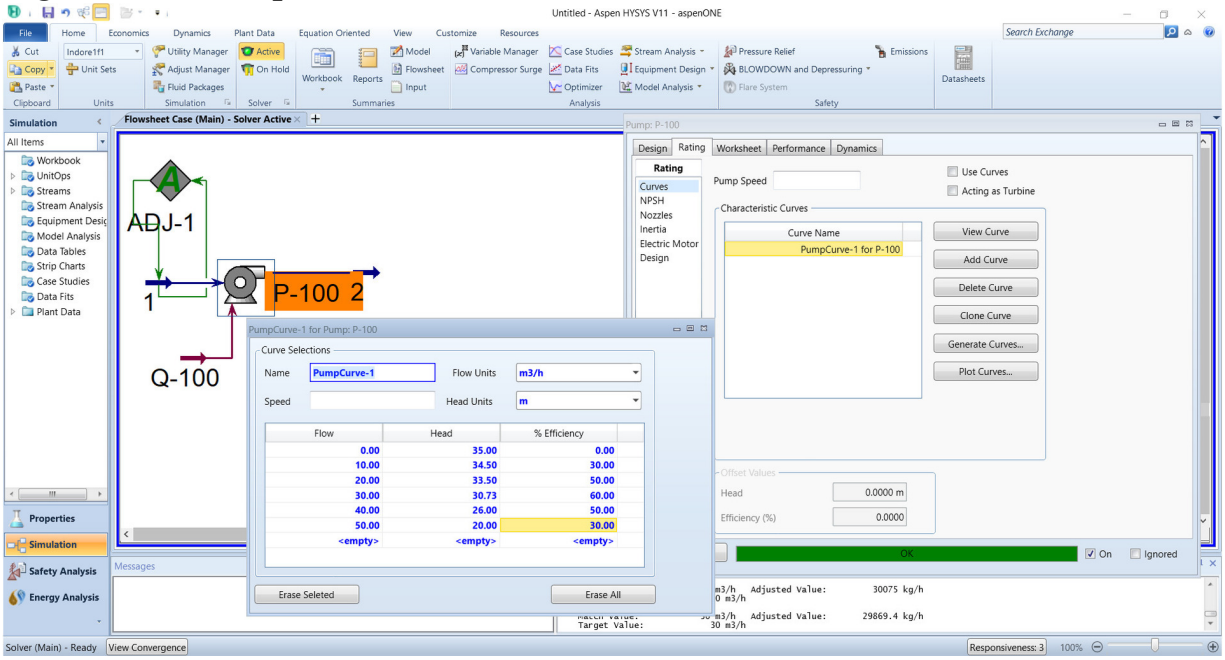
Select “Rating\NPSH” and check the “NPSH activate” box. HYSYS calculates $NPSH_{available}$ as 10 mlc.

Figure 6.13: Pump $NPSH_A$



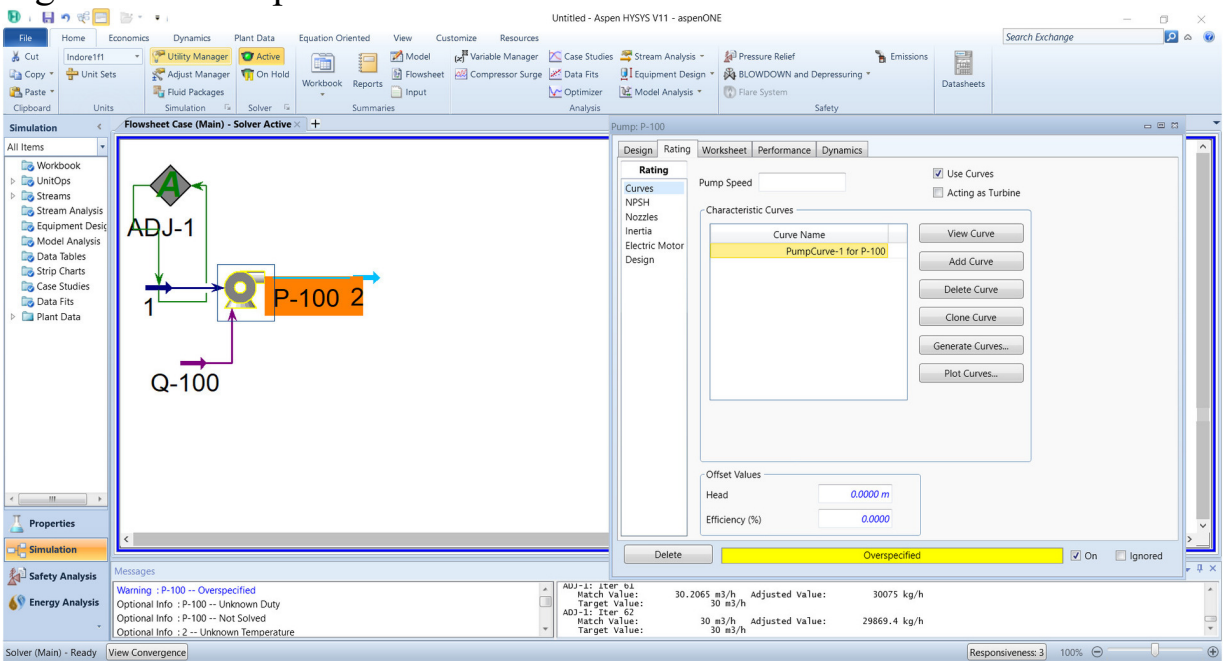
We can also provide pump curve data (H v/s Q and Efficiency v/s Q). Select "Rating\Curves" tab. "Add Curve" and provide the flow rate, head, and efficiency. This data should come from the pump vendor catalog.

Figure 6.14: Pump curve



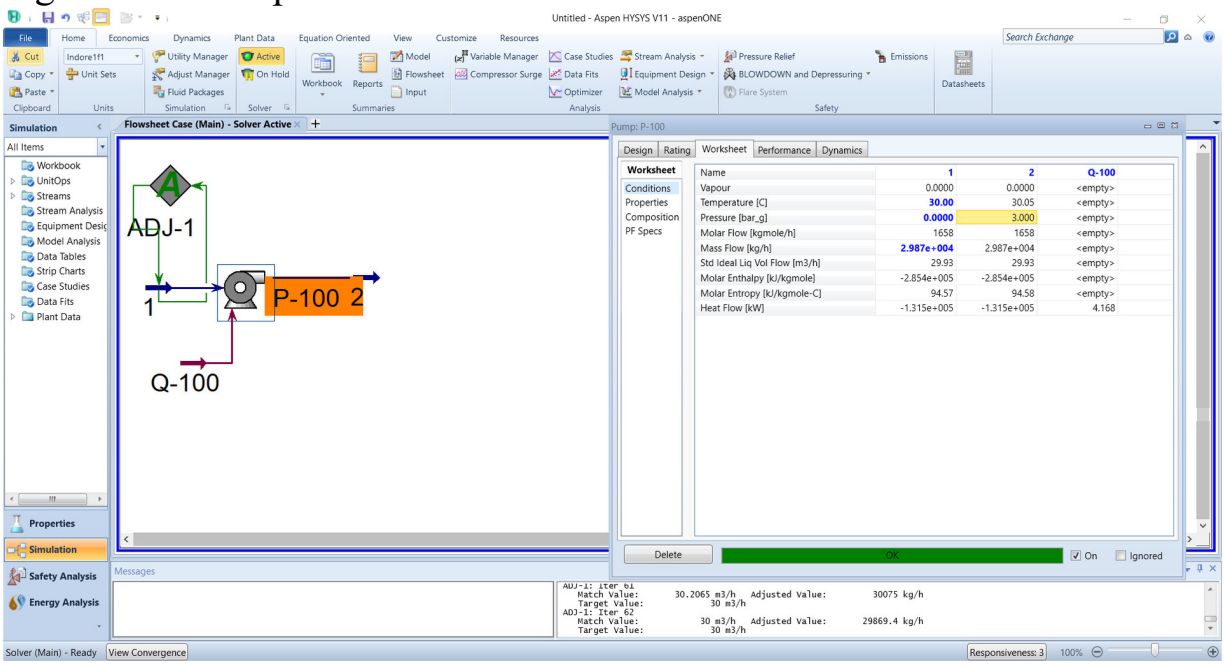
Check the “Use Curve” box, so that the provided pump curve data is used in the model. But HYSYS points out that it is “Overspecified”. That means we have given more information than required.

Figure 6.15: Pump curve



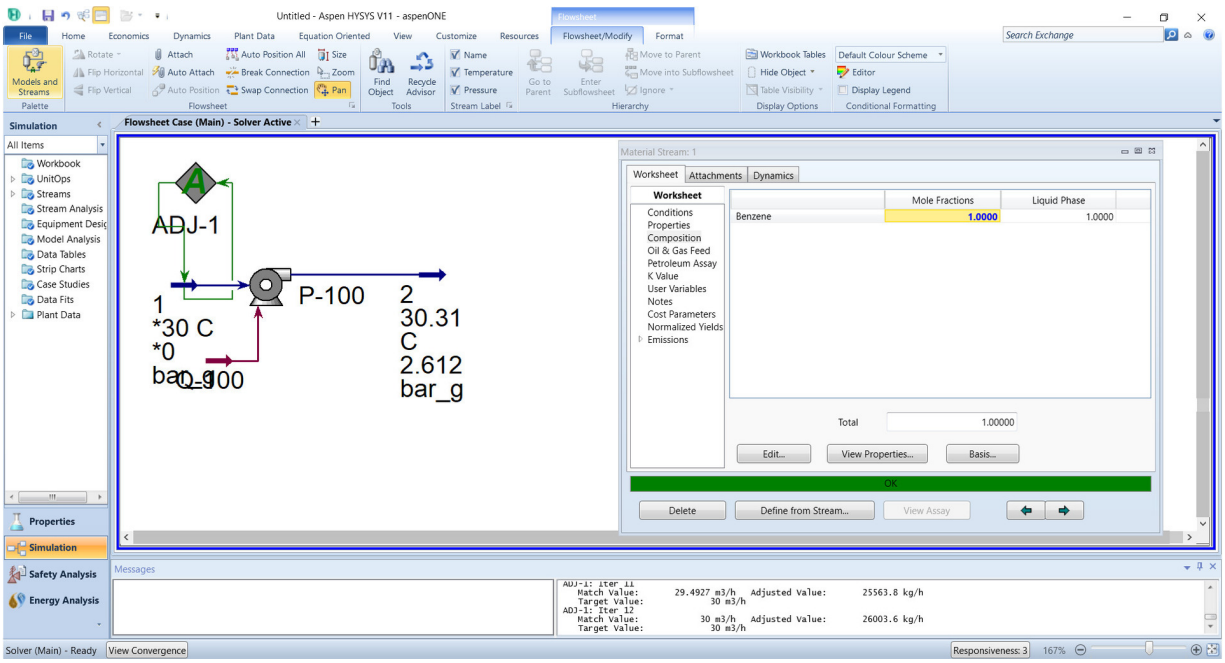
It may be noted that we have earlier provided pump efficiency (60%) and discharge pressure (3 barg) as input parameters. We will now delete these 2 input parameters. Also, provide a table for the pump as shown below.

Figure 6.16: Pump



Now if we use the same pump (with same pump characteristic curves), but change the liquid from water to benzene, then let's see what would be the effect on pump discharge pressure and $NPSH_{available}$ values. Use NRTL as a fluid package.

Figure 6.17: Pump



It may be noted that discharge pressure and pump shaft power has decreased. Can you justify these results?

Exercises

a. Ms. Sharvari has been given an assignment to present types of pumps. Please help her in her task.
b. Mr. Chirag works for an operating company as a production engineer. During his round, he observed that a cooling water pump was having abnormal noise and vibrations. What are all the possible causes for such a problem?
c. Mr. Hrishikesh was asked to collect 3 pump curves from different pump manufacturers and explain the same to his boss. Can you help him?

d. Ms. Sneha is simulating a fire water pump with capacity 75 m^3 /h and differential head as 45 mlc. If pump and motor efficiencies are 70% and 90% respectively, then what motor she would select?
e. Mr. Shivanand is production head and asks Ms. Sneha if the fire water pump can be used for benzene transfer temporarily due to some plant upset. What are the concerns Mr. Shivanand would have before using a fire water pump for benzene? Benzene flow rate and differential head required are 60 m3/h and 40 mlc respectively. Will it work as per the process? If not, how would you make it work?
f. Mr. Kalpesh has taken over from Mr. Hrishikesh and was assigned to use 3 pump curve data in the simulation model. Please help him.

CHAPTER 7: CONTROL VALVE SIZING USING ASPEN HYSYS

Three inter-dependent parameters work in tandem whenever there is a flow across a system; viz. Flow rate (Q), Pressure drop (ΔP), and flow area (A).
It can be visualized in the below table.

Table 7.1: Interdependency of A, ΔP and Q

Flow area	**Pressure drop**	**Flow rate**
Constant	Increases	Increases
Constant	Decreases	Decreases
Increases	Constant	Increases
Decreases	Constant	Decreases
Increases	Decreases	Constant
Decreases	Increases	Constant

It can be read as, e. g. for the same cross-sectional area available for flow, the flow rate will increase with the increase in pressure drop across it. This is the basic principle of fluid dynamics. A control valve also works on the same principle of A, ΔP, and Q inter-dependency.

Choked flow is a limiting condition where the mass flow doesn't increase with a further decrease in the downstream pressure environment for a fixed upstream pressure and temperature. The choked flow of gases is useful in many engineering applications because the mass flow rate is independent of the downstream pressure, and depends only on the temperature and pressure and hence the density of the gas on the upstream side of the restriction. Under choked conditions, valves and calibrated orifice plates can be used to produce a desired mass flow rate. A common example can be given as the fuel gas supplied to pilot burners at the flare stack top is

tapped from a pressurized fuel gas header. This tapping usually has a restriction orifice, that leads to a pressure drop exceeding 50% of the inlet pressure. It ensures choked flow, meaning unless there is any fluctuation in fuel gas header pressure, flow to pilot burners would remain the same. Orifice shall prove to be a cheaper option in such cases as compared to the flow control valve. The 50% figure mentioned above is easy to remember. Each gas has a different value of Ideal Gas Critical Flow Pressure Ratio, like Methane is 0.54, Steam is 0.54 and Air is 0.53.

If the fluid is a liquid, a different type of limiting condition (choked flow) occurs when the restriction in the valve causes a decrease of the liquid pressure beyond the restriction to below that of the liquid's vapor pressure at operating temperature. At that point, the liquid will partially flash into bubbles of vapor and the subsequent collapse of the bubbles causes cavitation. Cavitation is quite noisy and can be sufficiently violent to physically damage valves, pipes and associated equipment. In effect, the vapor bubble formation in the restriction prevents the flow from increasing any further. If a control valve has such a cavitation problem, then the vendor may choose another valve with anti-cavitation trim (that has high F_L value or liquid pressure recovery factor). The control valve with a high value of F_L is costlier.

What is a control valve?

- A control valve is a valve used to control fluid flow by varying the size of the flow passage as directed by a signal from a controller.
- This enables the direct control of flow rate and the consequential control of process quantities such as pressure, temperature, and liquid level.

What is a Cv?

Cv is a flow coefficient that provides the capacity of a valve.

Where;

- Q : Flow rate (USgpm)
- SG : Specific gravity of the fluid (for water =1)
- ΔP : Pressure drop across the valve (psi)

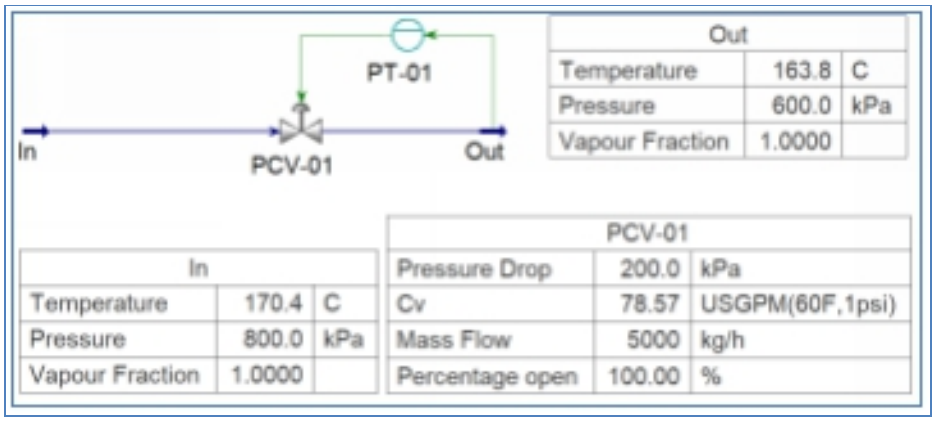
In more practical terms, Cv is the volume (in US gallons) of water at 60° F that will flow per minute through a valve with a pressure drop of 1 psi across the valve.
If a valve has a Cv of 20 that means we can pass 20 gallons of water per minute with a one-pound pressure drop.

Let's consider an example given in figure 1.

Saturated steam at 800 kPa enters the pressure control valve (PCV-01). PCV-01 maintains a downstream pressure of 600 kPa. The calculated capacity (Cv) of such a control valve shall be 78.57. When the valve is 100% open then for 200 kPa pressure drop, it shall allow 5000 kg/h steam to pass through.

If we use the same valve and decrease the pressure drop across from 200 kPa to 100 kPa, what happens to the flow rate using guidelines given in table 7.1. Less pressure drop means less flow, right?

Figure 7.1: Control valve

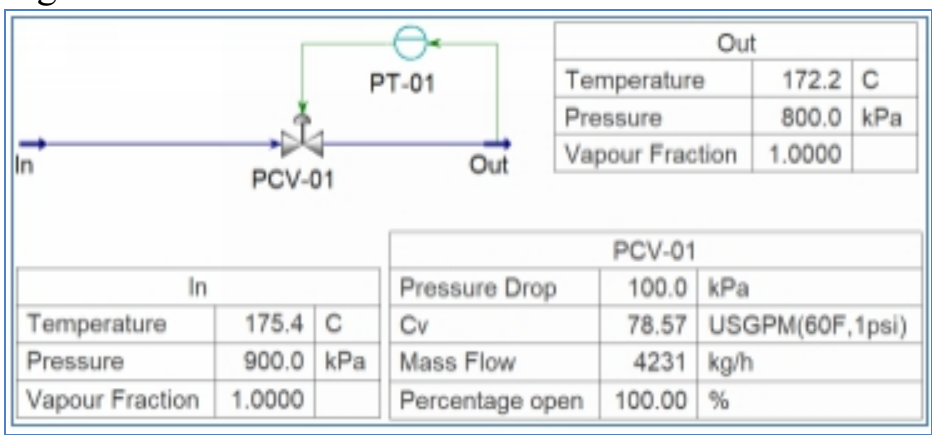


Out		
Temperature	163.8	C
Pressure	600.0	kPa
Vapour Fraction	1.0000	

In		
Temperature	170.4	C
Pressure	800.0	kPa
Vapour Fraction	1.0000	

PCV-01		
Pressure Drop	200.0	kPa
Cv	78.57	USGPM(60F,1psi)
Mass Flow	5000	kg/h
Percentage open	100.00	%

Figure 7.2: Control valve



Out		
Temperature	172.2	C
Pressure	800.0	kPa
Vapour Fraction	1.0000	

In		
Temperature	175.4	C
Pressure	900.0	kPa
Vapour Fraction	1.0000	

PCV-01		
Pressure Drop	100.0	kPa
Cv	78.57	USGPM(60F,1psi)
Mass Flow	4231	kg/h
Percentage open	100.00	%

Control Valve Sizing Criteria

- Valve shall be sized for normal operating flow at normal pressure drop. The valve selected shall have the capability to pass 1.3 times the normal flow at the indicated differential pressure and 1.1 times the maximum flow at the indicated differential pressure. The larger valve shall be selected.
- When only the maximum flow is specified, the valve shall be sized for 110 % of maximum flow (maximum expected flow rates at no more than 90% of the fully open position).
- Normal flow rates at 50% to 80% of the maximum opening of the valve and minimum flow rate no less than 10% open position.
- Butterfly valves shall be selected such that the disc opening is not greater than 60° at maximum flow and 30° for minimum flow. 90° is permissible for on/off service only.
- Control valve body size shall not be less than two-line sizes smaller than line size. Often, the valve body shall be one size less than the line size.
- Differential pressure for the valve shall not be less than 10 psi. If the system cannot afford 10 psi or more drop through the valve, a butterfly, eccentric disc or ball valve should be considered.

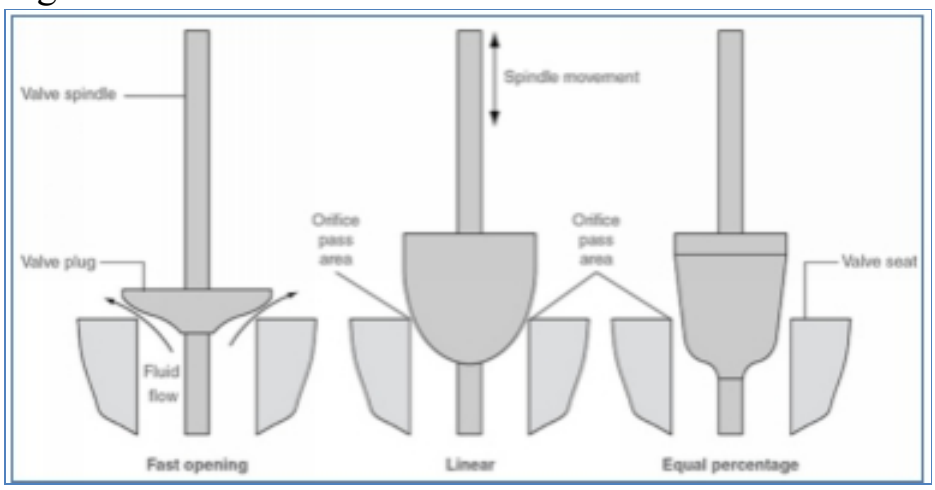
Valve Sizing Criteria

- Non-Critical Service - Outlet velocity < 8 m/s
- Non-Critical, corrosive Service - Outlet velocity < 6 m/s
- Partial Cavitation Service - Outlet velocity < 5 m/s

- Maximum and Severe Cavitation Service - Outlet velocity < 3 m/s
- Flashing Service - Inlet velocity < 5 m/s
- Flashing Service - Outlet velocity < 60 m/s
- Gas service (continuous) – Mach No. < 0.3
- Gas service (intermittent) – Mach No. < 0.7

Common flow characteristic curves

Figure 7.3: Trim characteristics

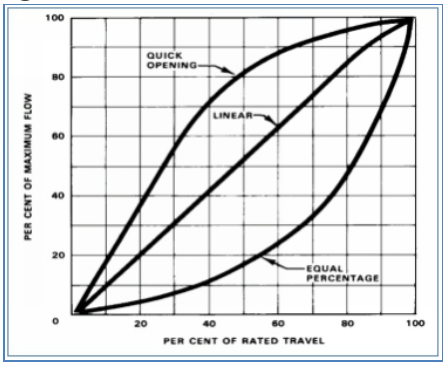


The above figure provides the tentative shapes of the valve plug or trim. Different shape of the valve plug decides the flow characteristic w.r.to valve opening. For example, if the valve characteristic is linear and for 100% valve opening the flow rate is 1000 kg/h, if you throttle the valve to 60%, then the flow rate will reduce to 600 kg/h, provided pressure drop in both the cases remain same.

- Linear**:**
    - Flow capacity increases linearly with valve travel.
    - Mostly used in liquid level or flow loops.
    - Used in systems where the pressure drop across the valve is expected to remain fairly constant (i.e. steady-state systems).
- Equal percentage:
    - Flow capacity increases exponentially with valve trim travel; equal increments of valve travel produce equal percentage changes in the existing Cv.
    - Most commonly used control valve
    - Used in processes where large changes in pressure drop are expected.
    - Used in temperature and pressure control loops.
- Quick Opening characteristic:
    - An inherent flow characteristic in which a maximum flow coefficient is achieved with minimal closure member travel
    - Used for frequent on-off service.
    - Used for processes where the "instantly" large flow is needed (i.e. safety systems or cooling water systems).

The typical trim characteristic can be plotted as below.

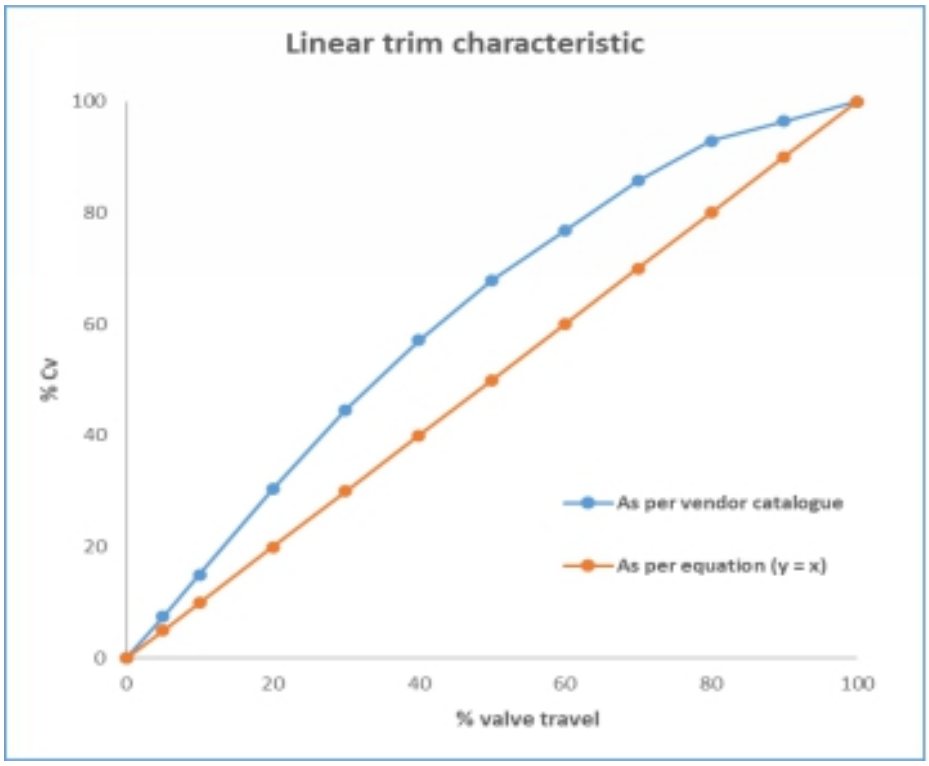
Figure 7.4: Trim characteristics



- Linear trim would follow the equation $y = x$.
- Equal percentage trim would follow the equation $y = x^3$.
- The quick opening trim would follow the equation $y = x^{0.5}$.

It's advisable to use the above equations when you are working on concept or FEED level study when the control valve vendor catalog is not available. However, if control valve sizing is carried out in detail engineering stage, where EPC contractor will buy the control valve, then it's mandatory that you use vendor catalog and do not rely on the equations above. It may be noted that mostly the actual valve behavior is different from the calculated one. One such example is plotted below.

Figure 7.5: Trim characteristics curve comparison



Control Valve selection process

- Determine service conditions.
- P1, ΔP, Q, T1, Fluid Properties, etc.
- Calculate preliminary cv required.
- Select the trim type.
- Select the valve body and trim size.

Self-acting regulators

Self-acting regulators shall only be used in services that meet all the following requirements:
Fluid services that are all the following:

- Clean.
- Not very toxic.
- Below its auto-ignition temperature.
- Permanent installations that require no operator intervention (e.g., changing of the setting).

All components downstream of the pressure-reducing regulator shall meet one of the following:

- Rated for the upper design pressure of the upstream piping, or
- Provided with overpressure protection.

Adjustments to the setpoint of self-acting regulators shall only be possible with a tool such as a wrench or spanner and not by hand.

Types of self-acting regulators

Pressure-reducing regulators

Pressure reducing regulators should be used to control the pressure downstream of the regulator. These valves are failed to open. The actuator casing is connected to the outlet pressure, so in case of control failure, the downstream piping and actuator casing will be exposed to the inlet pressure. The non-pressurized side of the actuator normally has a breathing connection which, in case of diaphragm failure will pass process fluid to the atmosphere.

Back-pressure regulators

Back-pressure regulators should be used to control the pressure upstream of the regulator. These valves are failed to close. The actuator casing is exposed to the upper design pressure of the upstream piping. The non-pressurized side of the actuator normally has a breathing connection which, in case of diaphragm failure, will pass process fluid to the atmosphere.

Differential-pressure regulators

The pressure retaining components of the differential-pressure regulator shall be rated for the upper design pressure of the upstream and downstream piping, whichever is the highest.

Tank blanketing regulators

Tank blanketing regulators are used to feed blanket gas into a tank when the pressure in the tank drops, for instance when the level is dropping. External pressure sensing lines of tank blanketing regulators should be installed with minimal bends and constant slope to allow for drainage back into the tank. In fouling or dirty applications, the provision of a constant purge of gas into the tank should be considered.

Temperature regulators

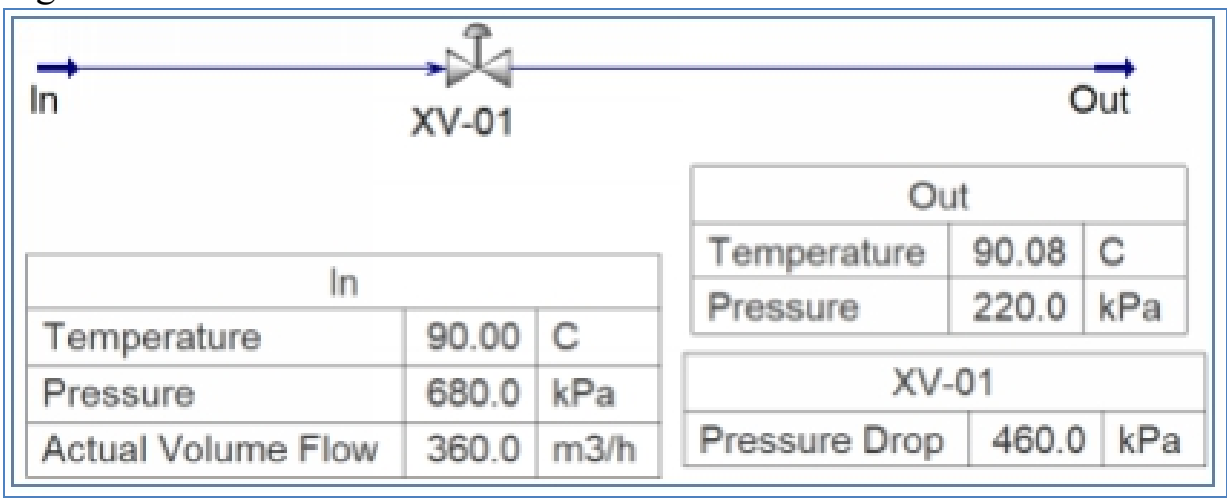
Self-acting temperature regulators shall only be considered for heating applications where utilities (e.g., instrument air or gas) are not available.

Simulation

In the below example, 360 m^3 /h of water is passed through an ON/OFF valve (XV-01) at 90 ° C. Inlet pressure is 680 kPa. A pressure drop value of 460 kPa is provided to XV-01.

Calculate the appropriate size of XV-01.

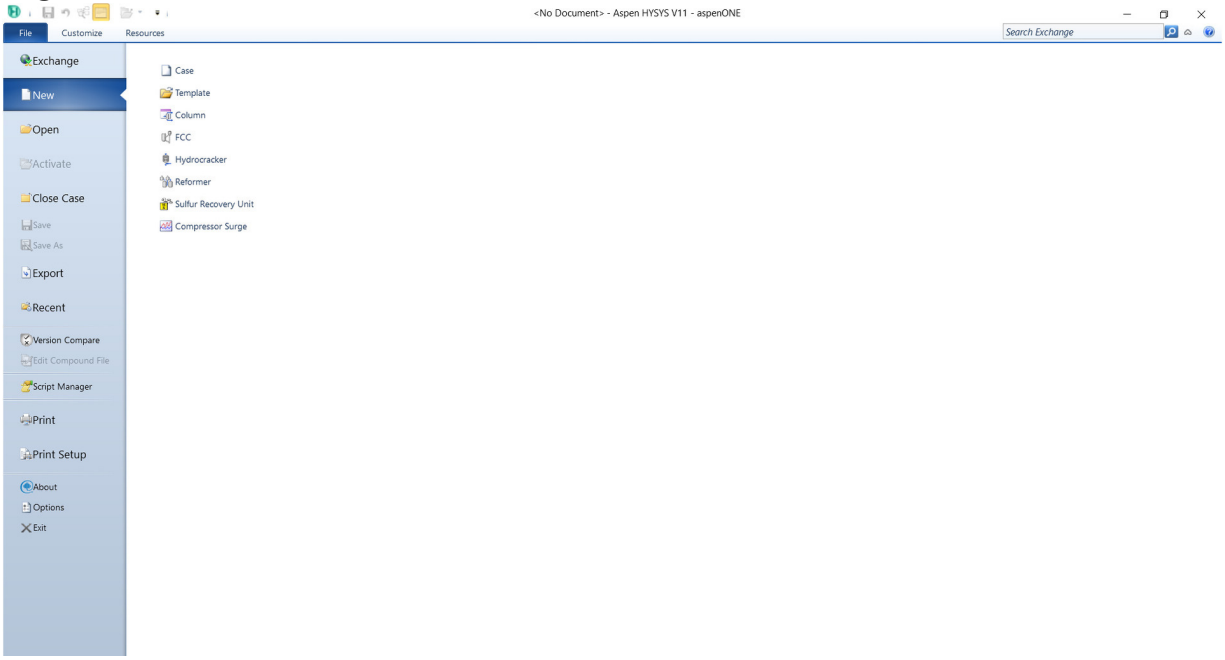
Figure 7.6: Control valve



Step by step procedure for simulating this system in Aspen HYSYS V11 is provided below;

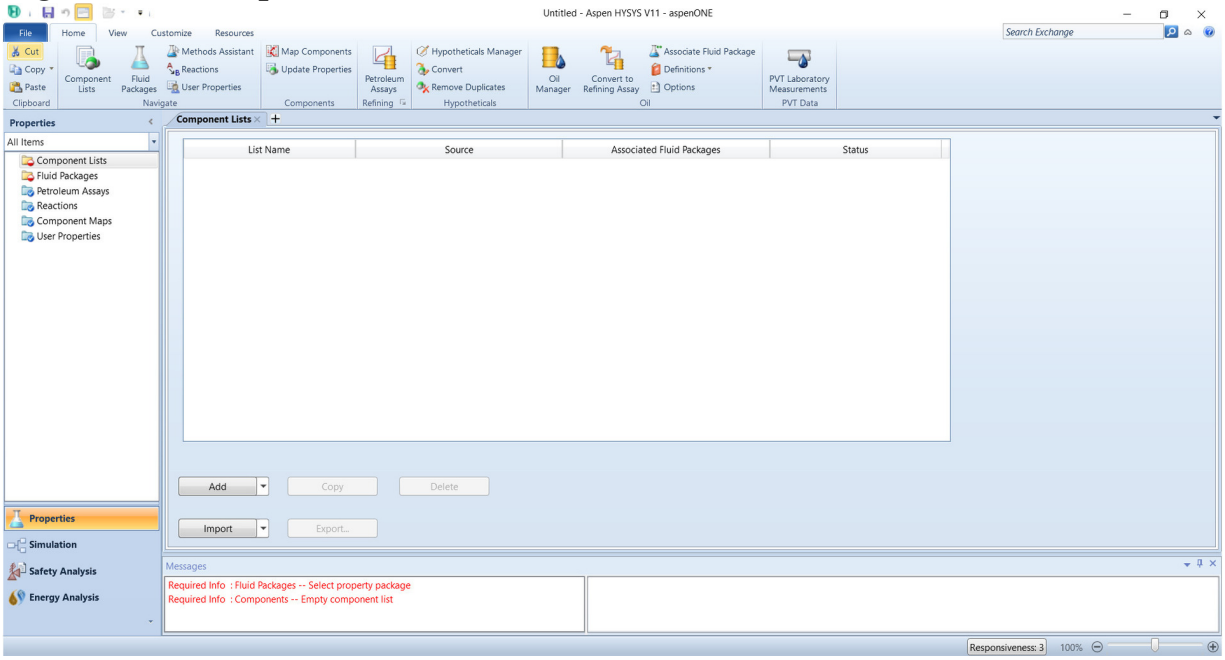
Create a new case in Aspen HYSYS.

Figure 7.7: New case



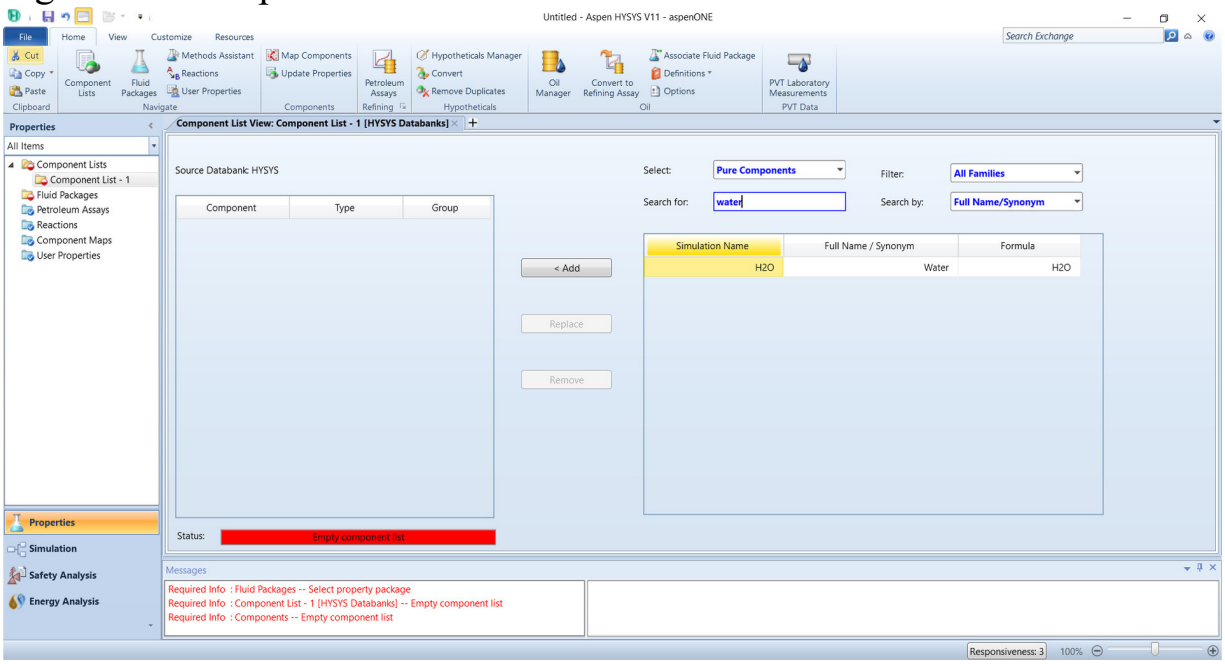
“Add” the “Component Lists”.

Figure 7.8: Component list



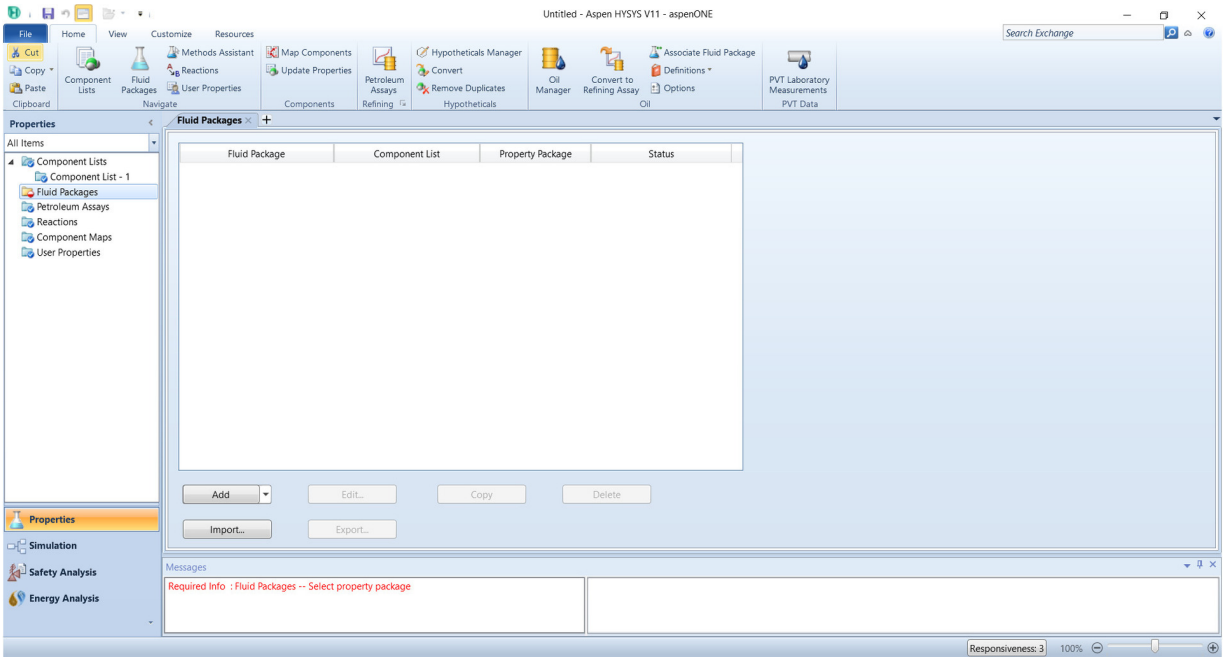
“Search for” Water as the “Pure Components” and “Add”. There are a total of 5758 pure components in the HYSYS V11 library. The Full Name/Synonym option makes finding components the easiest. To enter components by HYSYS’s simulation names, or by chemical composition, select “Sim Name” or “Formula” respectively.

Figure 7.9: Component list



In the “Fluid Packages”, click on the “Add” button to specify a fluid package. Doing so will bring you to a list of all the different equations of state HYSYS uses. The simplest fluid package, you might be aware of the Antoine equation for calculating the vapor pressure of a fluid.

Figure 7.10: Fluid package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File
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Component Lists
Fluid Packages
Methods Assistant
Reactions
User Properties
Navigate
Map Components
Update Properties
Components
Petroleum Assays
Refining
Hypotheticals Manager
Convert
Remove Duplicates
Hypotheticals
Oil Manager
Convert to Refining Assay
Associate Fluid Package
Definitions
Options
Oil
PVT Laboratory Measurements
PVT Data
Properties
All Items
Component Lists
Component List - 1
Fluid Packages
Petroleum Assays
Reactions
Component Maps
User Properties
Fluid Packages
Fluid Package
Component List
Property Package
Status
Add
Edit...
Copy
Delete
Import...
Export...
Properties
Simulation
Safety Analysis
Energy Analysis
Messages
Required Info : Fluid Packages -- Select property package
Responsiveness: 3
100%

See the figure below for some general guidelines on picking the appropriate fluids packages.

Figure 7.11: Fluid package selection guideline

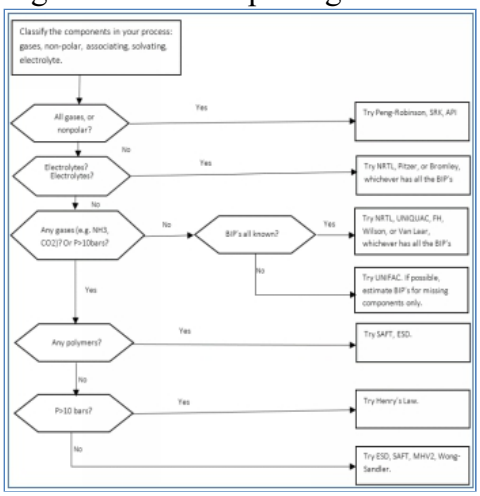


Table 7.2: Recommended property package

TYPE OF SYSTEM	RECOMMENDED PROPERTY METHOD
TEG Dehydration	PR
Sour Water	PR, Sour PR
Cryogenic Gas Processing	PR, PRSV
Air Separation	PR, PRSV
Ethylene Towers	Lee Kesler Plocker
Atm. Crude Towers	PR, PR Options, GS
Vacuum Towers	PR, PR Options, GS(<10 mmHg),Braun K10, Esso K
Hydrate Inhibition	PR
High H2 Systems	PR, ZJ or GS
Reservoir Systems	Steam Package, CS or GS
Chemical Systems	Activity Models, PRSV
HF Alkylation	PRSV, NRTL
H2O solubility in HC is important	Kabadi Danner
TEG Dehydration with Aromatics	PR
Systems with select gases and light HC	MBWR

Aspen HYSYS contains an assistant tool to help you in the selection for the suitable FP, called Methods Assistant.

Figure 7.12: Methods assistant

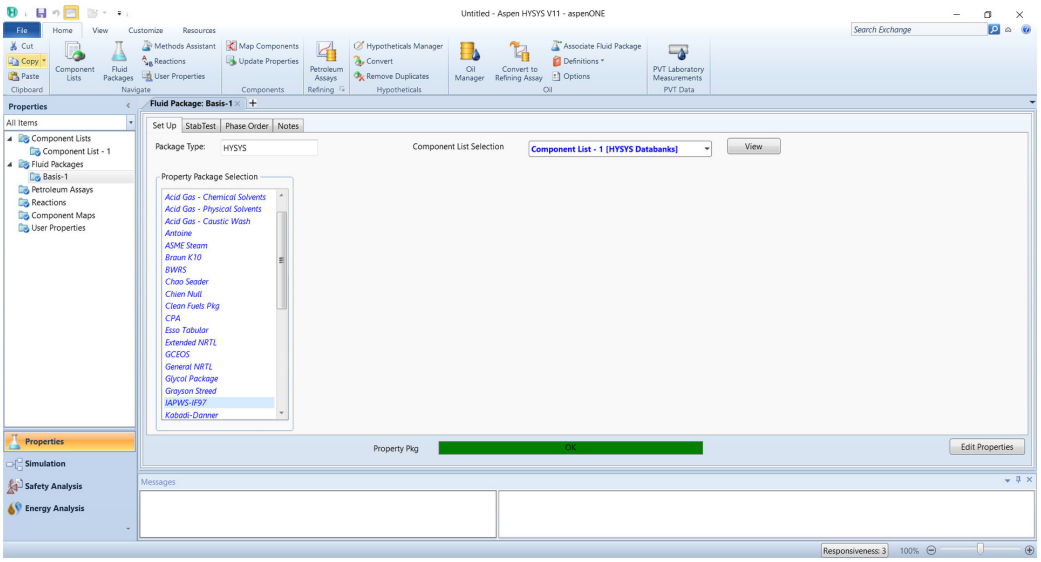


Figure 7.13: Methods assistant

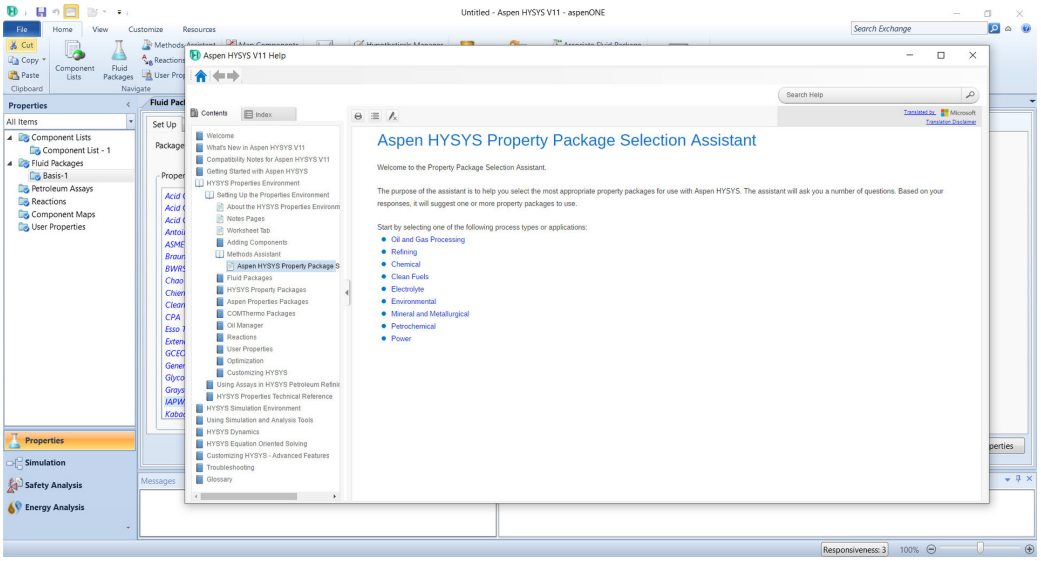
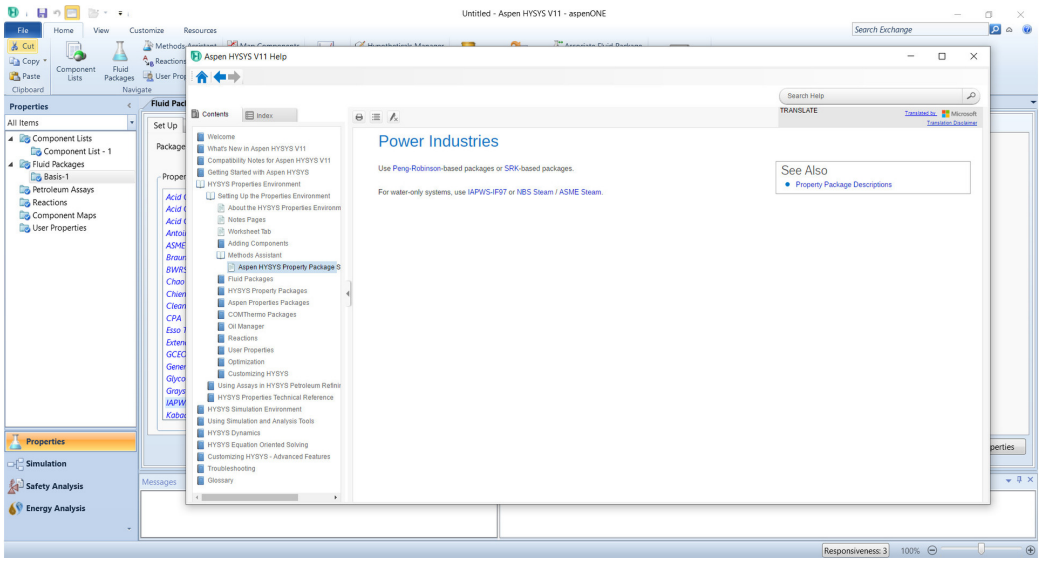


Figure 7.14: Methods assistant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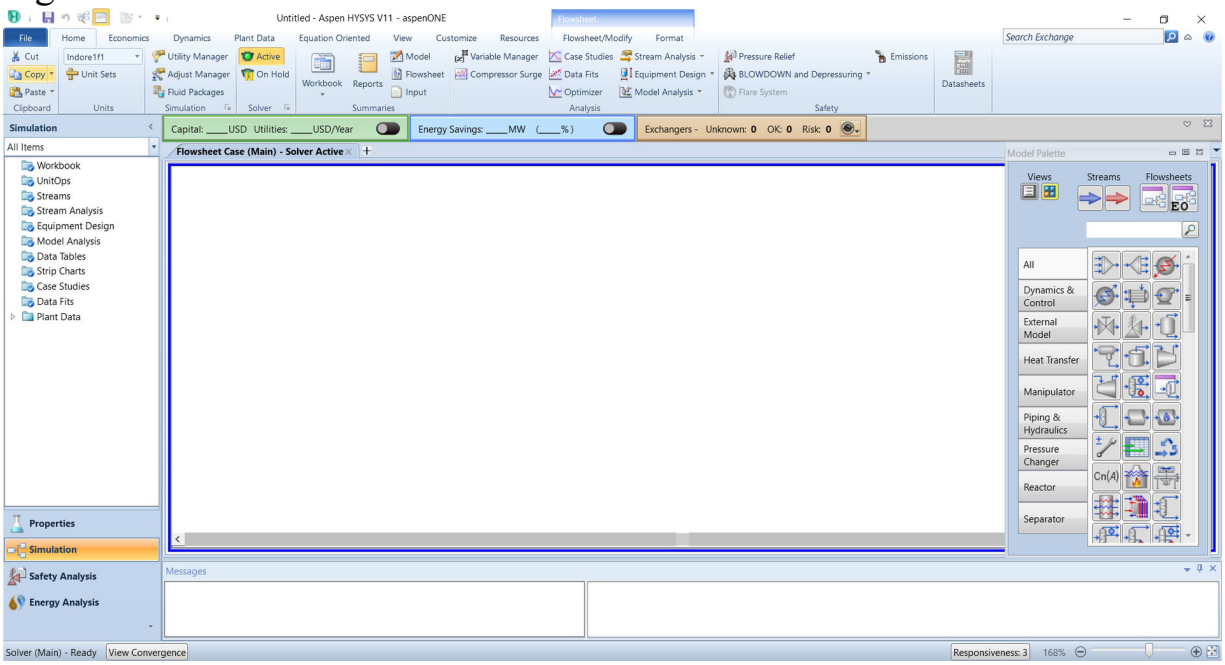
Aspen HYSYS V11 Help
Power Industries
Use Peng-Robinson-based packages or SRK-based packages.
For water-only systems, use IAPWS-IF97 or NBS Steam / ASME Steam.
See Also
Property Package Descriptions
Properties
Simulation
Safety Analysis
Energy Analysis

Using “Methods Assistant” one can find out that the recommended fluid package for the water-only system is IAPWS-IF97 or NBS-Steam or ASME Steam. We have selected IAPWS-IF97.

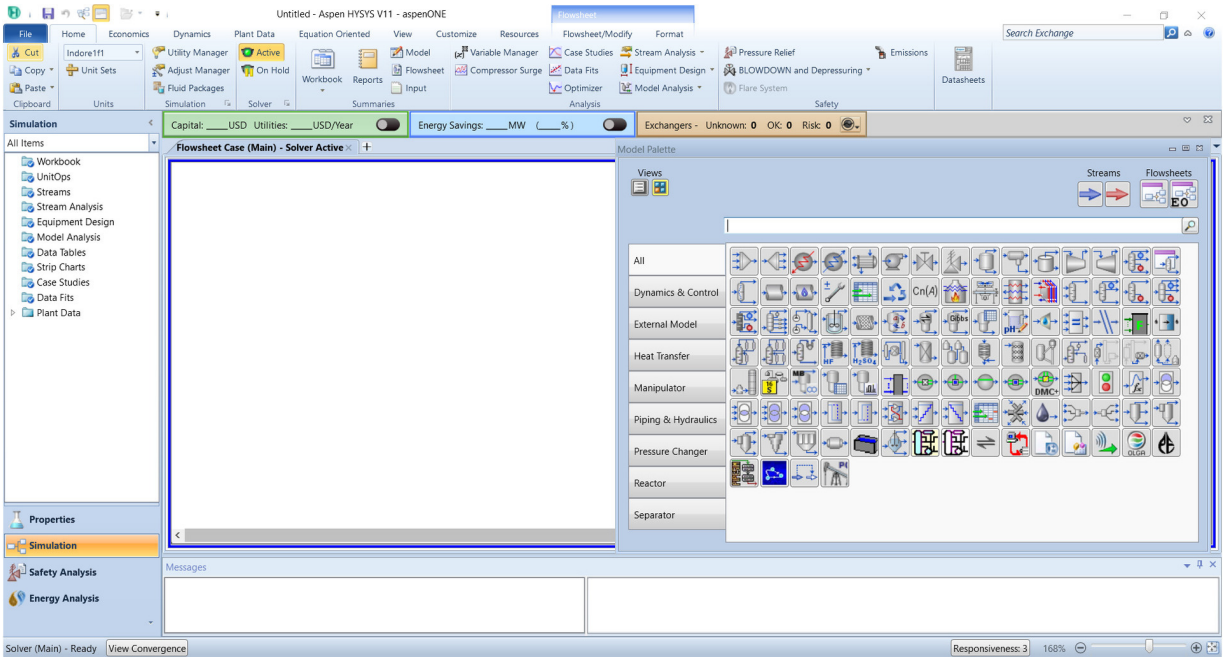
Enter the simulation environment by clicking the “Simulation” tab.

Figure 7.15: Simulation



“Model Palette” is used to insert any unit operation, material or energy stream, pump, compressor, control valve, distillation column, reactor, heat exchanger, etc.

Figure 7.16: Model palette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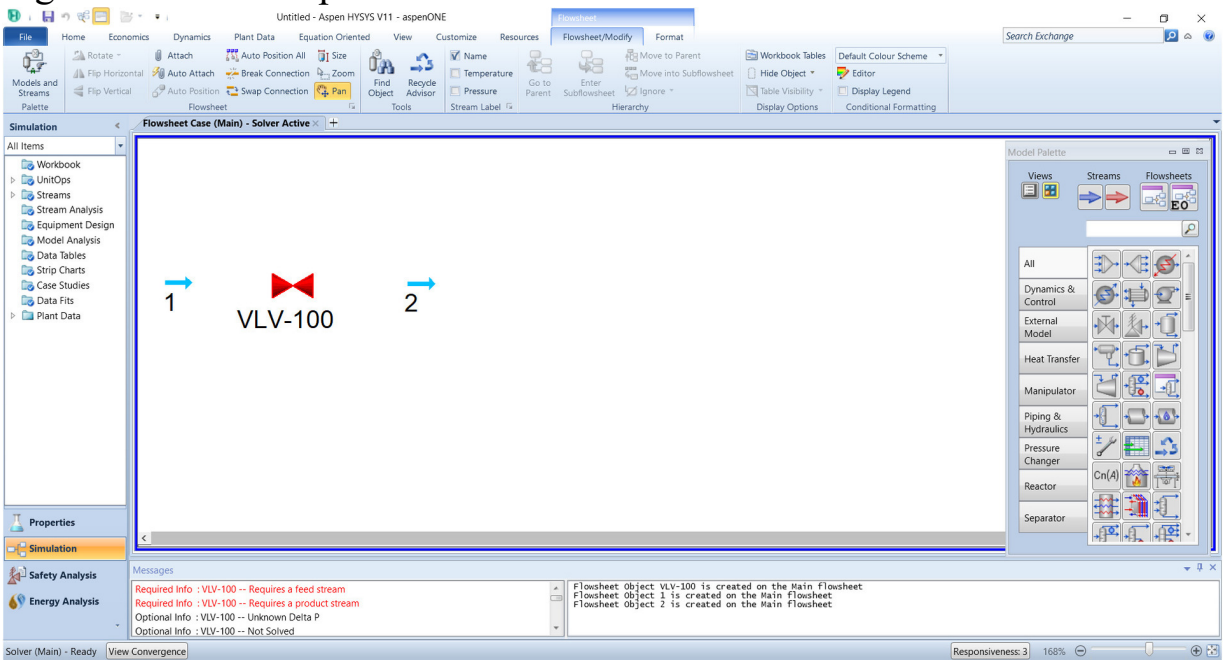
File
Home
Economics
Dynamics
Plant Data
Equation Oriented
View
Customize
Resources
Flowsheet
Flowsheet/Modify
Format
Search Exchange
Cut
Copy
Paste
Clipboard
Indore1f1
Unit Sets
Units
Utility Manager
Adjust Manager
Fluid Packages
Simulation
Active
On Hold
Solver
Workbook
Reports
Model
Flowsheet
Input
Summaries
Variable Manager
Compressor Surge
Case Studies
Data Fits
Optimizer
Stream Analysis
Equipment Design
Model Analysis
Analysis
Pressure Relief
BLOWDOWN and Depressuring
Flare System
Emissions
Safety
Datasheets
Simulation
All Items
Workbook
UnitOps
Streams
Stream Analysis
Equipment Design
Model Analysis
Data Tables
Strip Charts
Case Studies
Data Fits
Plant Data
Capital: ____USD Utilities: ____USD/Year
Energy Savings: ____MW (____%)
Exchangers - Unknown: 0 OK: 0 Risk: 0
Flowsheet Case (Main) - Solver Active
Model Palette
Views
Streams
Flowsheets
All
Dynamics & Control
External Model
Heat Transfer
Manipulator
Piping & Hydraulics
Pressure Changer
Reactor
Separator
Properties
Simulation
Safety Analysis
Energy Analysis
Messages
Solver (Main) - Ready
View Convergence
Responsiveness: 3
168%

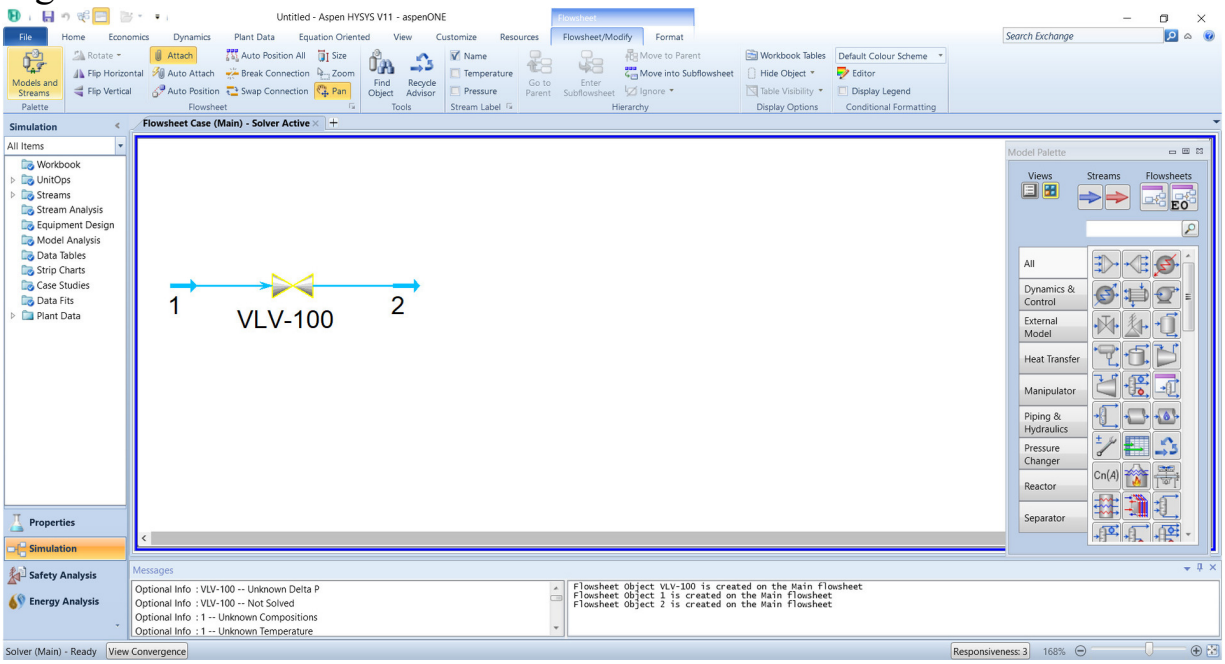
Insert 2 “Material Streams” and 1 “Control Valve”.

Figure 7.17: Model palette



Connect both the material streams to the control valve. Click the "Flowsheet/Modify" tab and use the "Attach" function to connect the streams.

Figure 7.18: Control valve



Double click on Material Stream "1" and provide temperature, pressure, flow rate and composition as per problem statement.

Figure 7.19: Material stream

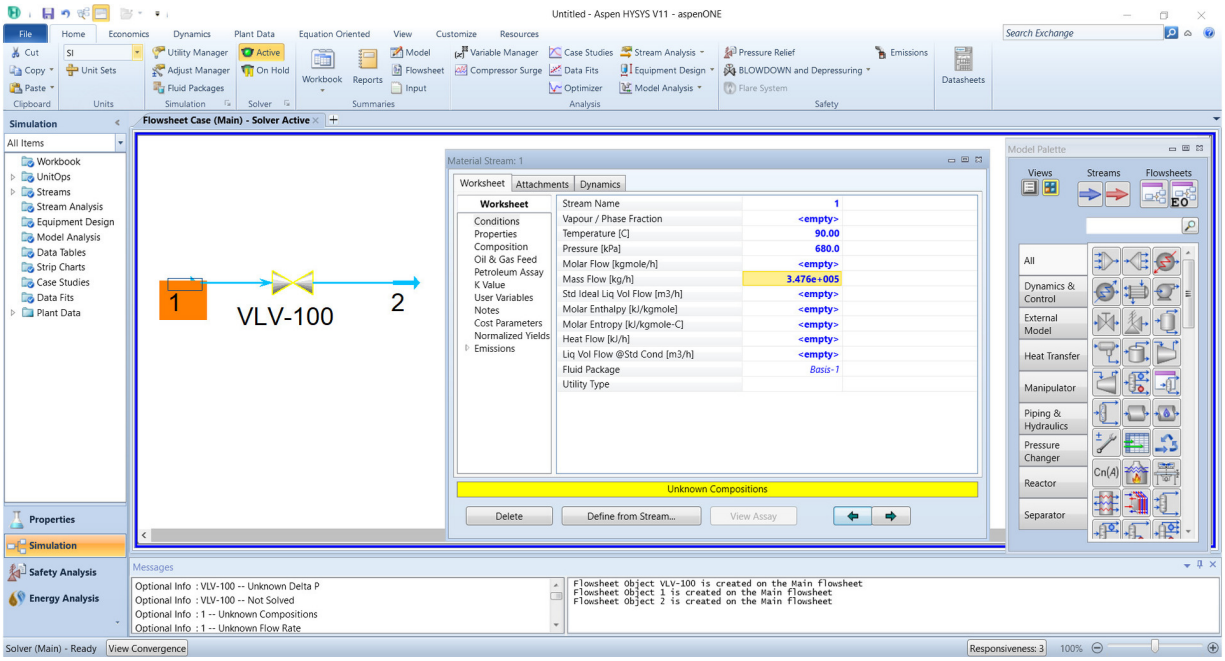
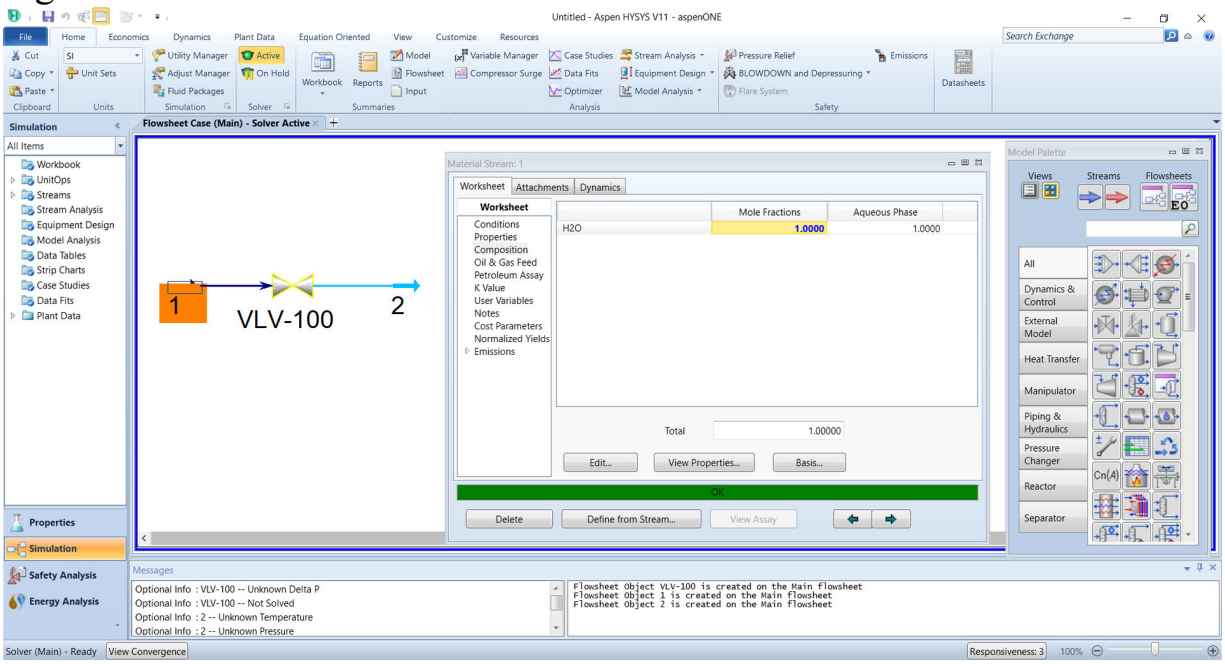
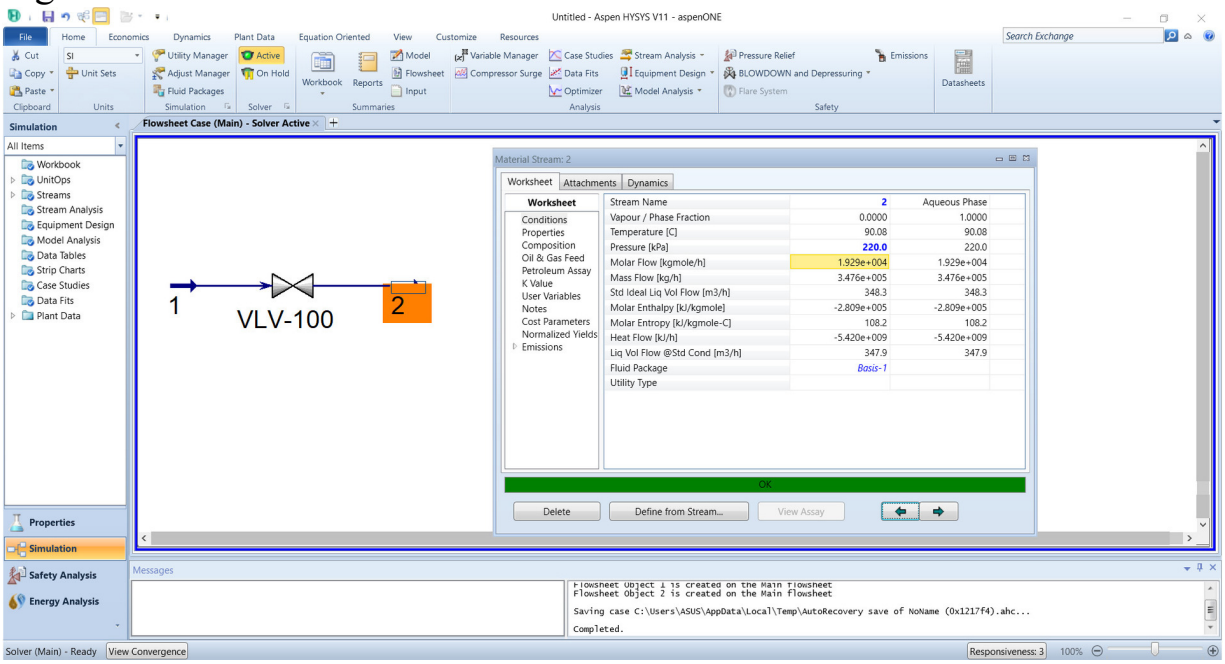


Figure 7.20: Material stream



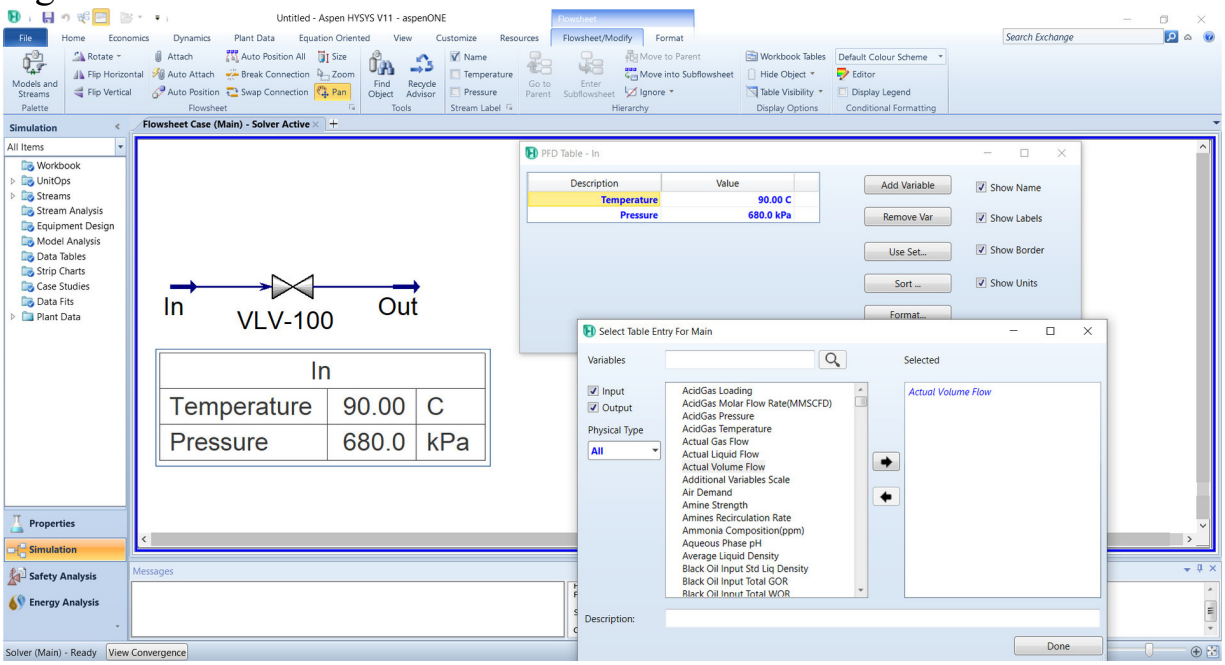
Similarly, provide pressure value of material stream 2. And change the tags of both the material streams and valve. Once all the minimum required input parameters have been provided to HYSYS, it shall calculate instantaneously. Material stream color shall change from cyan to blue and Valve color shall change from yellow to gray. Save the file as “Example 1.hsc”.

Figure 7.21: Material stream



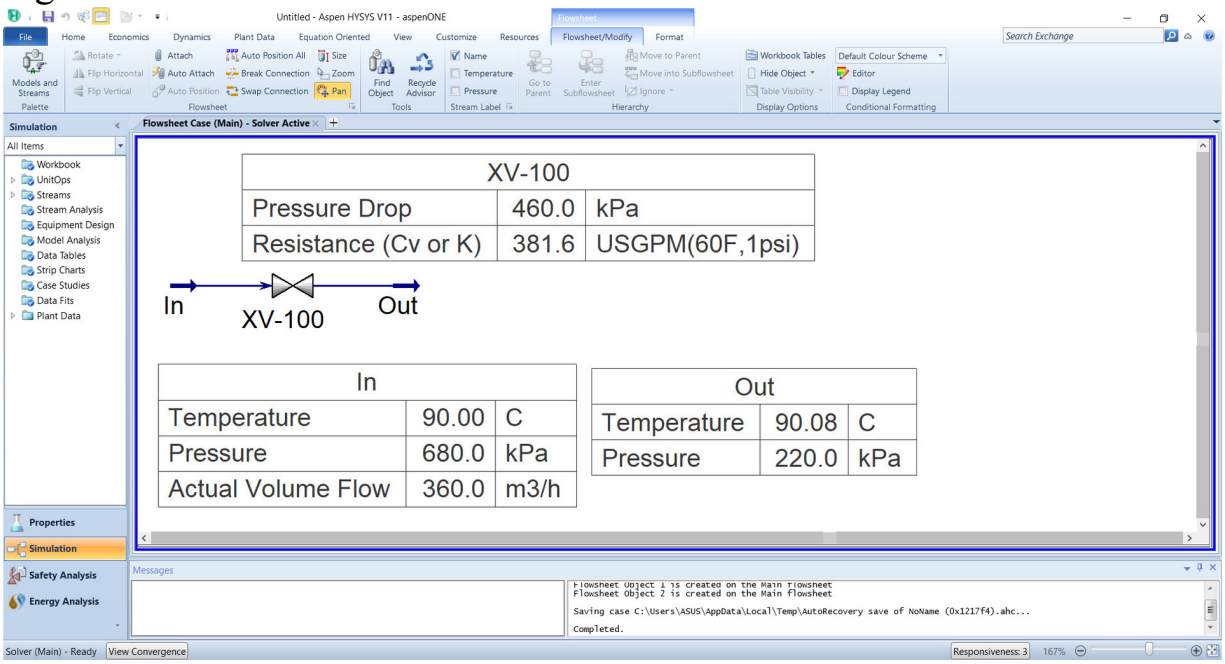
Right-click on stream “In” and select “Show Table”. Double click on the table and use “Add Variable” and “Remove Variable” to select and remove needed parameters as shown in the problem statement.

Figure 7.22: Show table



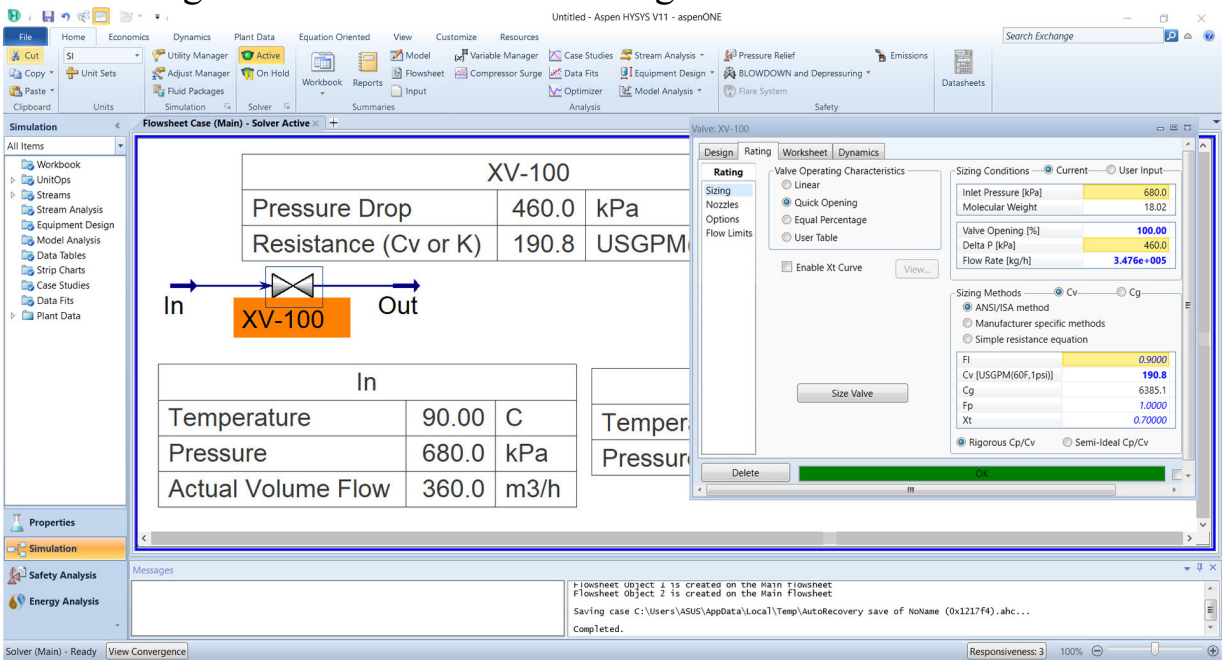
Do the same for other streams and valve as well.

Figure 7.23: Show table



As the system is converged, now we can proceed to control valve sizing. In the simulation, the sizing of the control valve is limited to Cv calculations. Double click on XV-01. Change "Valve Opening" to 100% as XV-100 is ON/OFF valve.

Figure 7.24: Control valve sizing



Control valve sizing procedure is provided in the standard ANSI/ISA 75.01.01. A few terms to ponder are as following;

- F_P : Piping geometry factor.

It can be calculated by the formula given in ANSI/ISA 75.01.01.

- F_L : Liquid pressure recovery factor of a control valve without attached fittings.

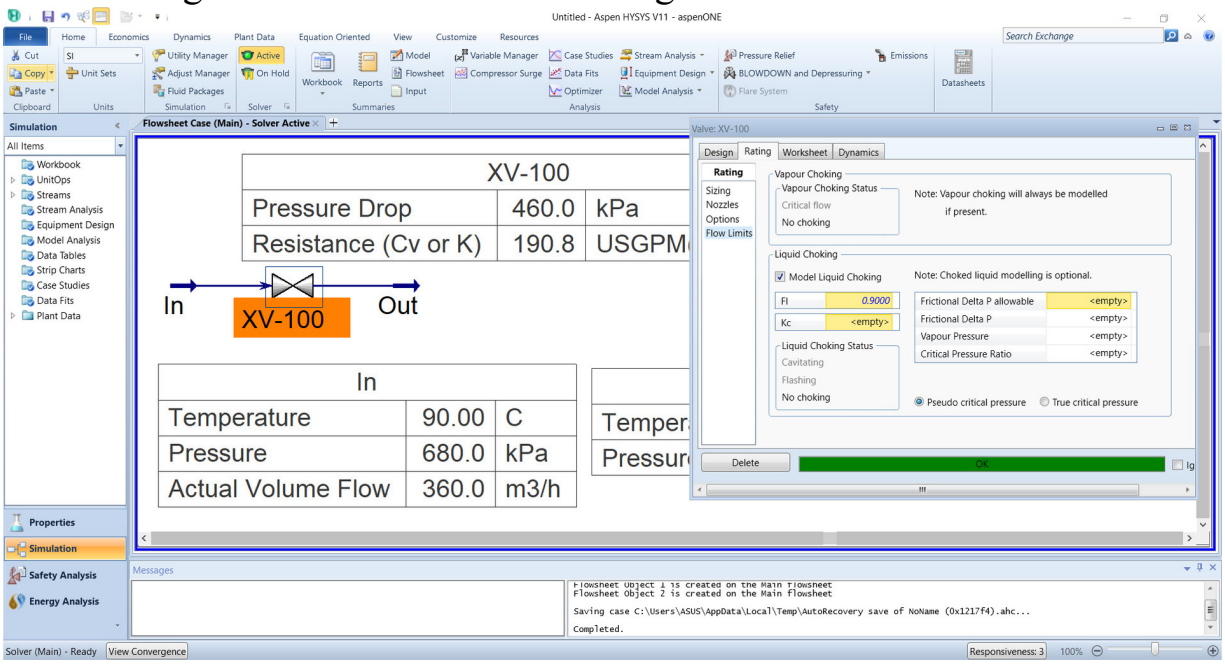
Typical values are provided in ANSI/ISA 75.01.01. Actual values shall be stated by the valve manufacturer. Less Fl value means the valve is more prone to choking.

- X_T : Pressure differential ratio factor of a control valve without attached fittings at choked flow.

Typical values are provided in ANSI/ISA 75.01.01. Actual values shall be stated by the valve manufacturer. Less Fl value means the valve is more prone to choking.

Also, select the “Rating” tab and check the “Model Liquid Choking” box. This will calculate if the valve has any choking problem or not. This works when you click “Size Valve” in the “Sizing” tab. In example 1, there is no choking observed.

Figure 7.25: Control valve sizing



Now let's go back to the "Rating" tab.

Cv calculated is 190.8. This would also mean that one can pass 190.8 gallons of water per minute with a one-pound pressure drop at 60 oF. Now we need to select an appropriate valve size for this application. Since it has been established that it is an ON/OFF valve, so trim characteristic suggested is a Quick opening. We shall refer control valve vendor catalog. One such table for Quick opening trim is given below;

Table 7.3: Cv table for quick opening trim

Valve Size (in.)	Trim Size (in.)	Stroke (in.)	Cv At Percent Open										
			5	10	20	30	40	50	60	70	80	90	100
0.5	0.5	0.75	0.4	0.74	1.4	4.5	3.1	3.8	3.8	4.1	4.4	4.6	4.8
0.75	0.72	0.75	0.9	1.7	3.0	4.8	6.3	7.2	7.2	7.3	7.4	7.4	7.5
1	0.81	0.75	0.5	1	3.0	6.0	8.3	9.3	9.9	10.3	10.4	10.5	10.5
1.5	1.25	1	2.5	4.9	8.9	14	20	25.5	28	29	29	29	30
2	1.62	1.5	4.1	8.2	15	28	39	46.4	47	49	49	49	50
3	2.62	2	11	22	39	65	91	111	125	127	127	128	129
4	3.5	2.5	19	38	69	111	154	187	213	218	222	225	225
6	5	3	38	76	143	222	296	362	420	463	465	466	466
8	6.25	4	59	117	222	360	479	592	681	693	706	716	726
10	8	4	96	191	367	544	755	840	980	1100	1130	1160	1180
12	9.5	4	123	246	495	738	940	1115	1280	1440	1570	1620	1670

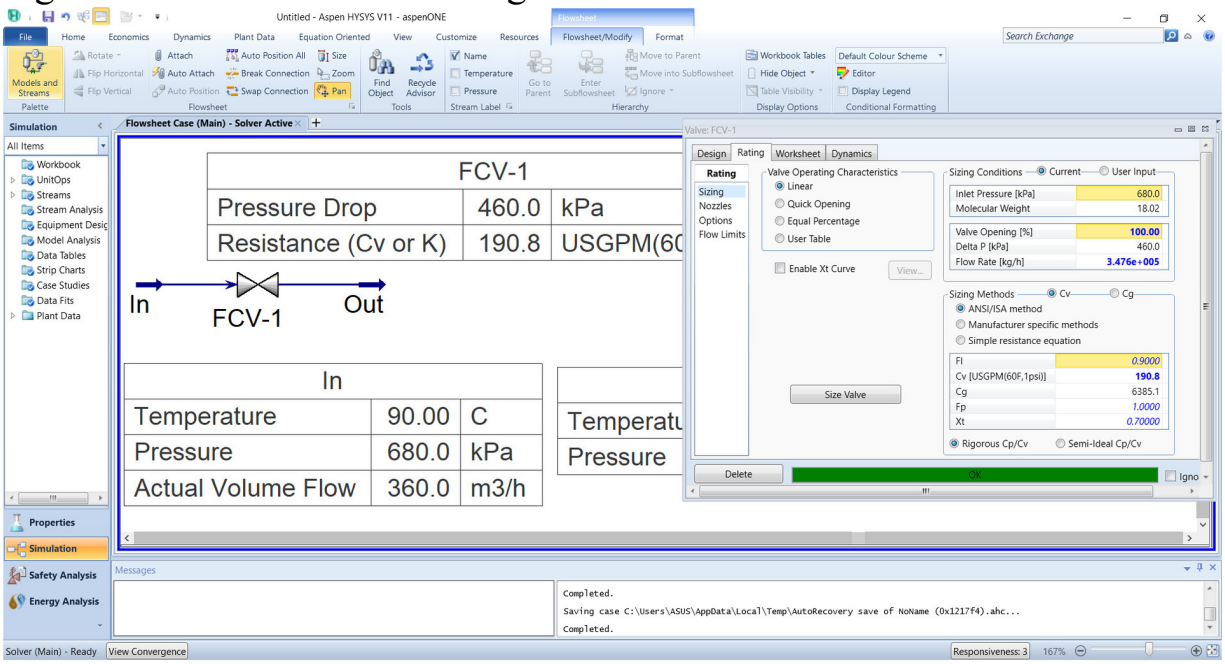
As the calculated Cv is 190.8, the adequate Cv as per the above table shall be 225. This falls under 4" valve size and 3.5" trim size. This is how the valve size is selected.

Now save this example as Example 2. In example 2, we will consider the same process parameters, but instead of the ON/OFF valve, XV-01 shall be a Flow control valve (FCV-01). Also, it is to be noted that the 360 m^3 /h (347608 kg/h) flow rate is the maximum or design flow rate. Most control valve vendors recommend operating range from 20% to 80%, meaning for minimum flow conditions, valve opening shall not below 20% and for

maximum flow conditions, valve opening shall not be more than 80%. This helps better control of the process parameter. For the flow control valve, the suggested
trim characteristic is Linear.

The model snapshot is given below;

Figure 7.26: Control valve sizing



The interpretation of the above results is as below;

Cv of 190.8 is required when the valve is at its maximum capacity, but in this case, the actual valve opening shall be 80%. Hence, we need to first get the vendor catalog for Linear trim and search for the Cv value equal or greater than 190.8 in the 80% valve opening column. (100% valve opening in the above snapshot is correct).

The below table is for the Linear trim control valve.

Table 7.4: Cv table for linear trim

Valve Size (in.)	Trim Size (in.)	Cv At Percent Open										
		5	10	20	30	40	50	60	70	80	90	100
1	0.81	0.95	1.9	3.6	5.1	6.4	7.2	8.0	8.5	9.0	9.3	9.6
	0.62	0.55	1.1	2.2	3.3	4.2	5.1	5.9	6.5	7.1	7.6	8.1
	0.38	0.21	0.41	0.82	1.2	1.6	2.0	2.4	2.8	3.2	3.5	4.0
	0.25A	0.09	0.18	0.37	0.55	0.73	0.91	1.1	1.4	1.5	1.6	1.9
	0.12A	0.024	0.047	0.093	0.14	0.18	0.23	0.28	0.32	0.37	0.42	0.47
2	1.62	3.8	7.6	15	20	25	29	32	35	36	37	38
	1.00	1.5	2.9	5.8	8.6	11	14	16	19	21	23	25
	0.72	0.95	1.9	3.8	5.7	7.6	9.4	11	13	15	16	18
	0.38	0.21	0.41	0.82	1.2	1.6	2.1	2.5	2.8	3.3	3.7	4.2
4	3.50	18	35	68	95	117	135	148	159	167	173	179
	1.62	3.9	7.8	15	23	31	38	45	54	60	67	75
6	3.50	18	36	71	104	136	165	192	216	239	259	278
	3.00	13	26	52	78	102	126	148	171	190	210	228
10	8.00	48	95	190	285	380	475	570	665	763	855	950
12	9.50	63	125	250	375	500	625	750	875	1010	1125	1250

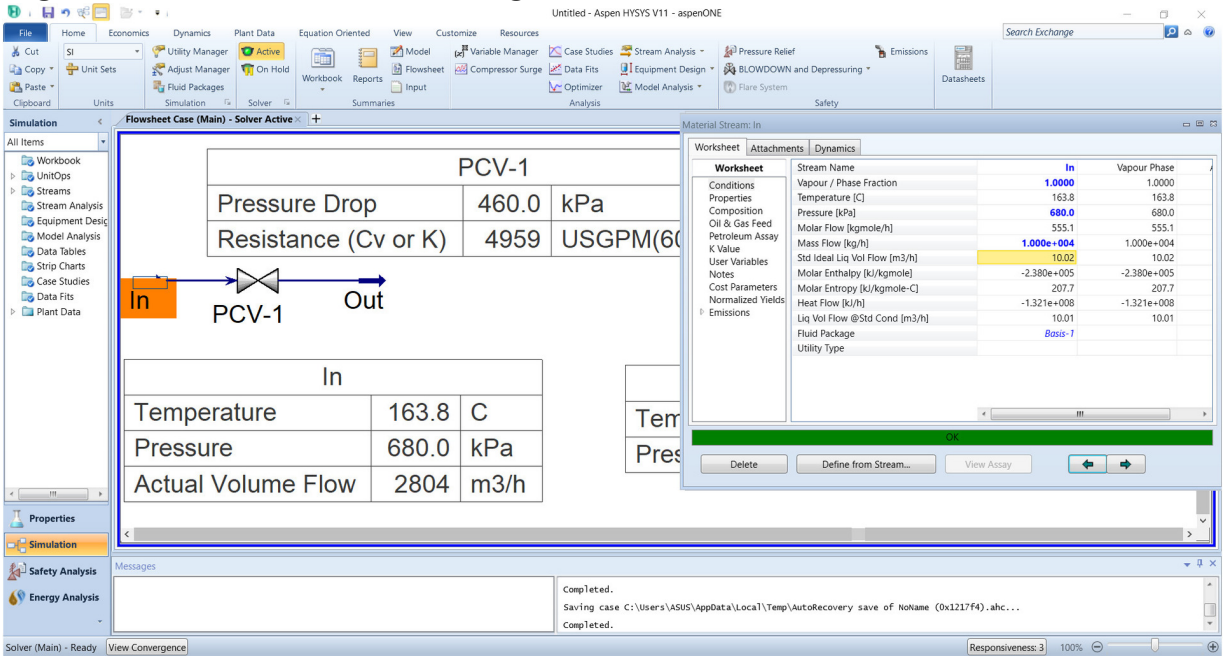
As per the above vendor catalog, FCV-01 size is 6” and trim size is 3”. Although Cv in 80% column is 190 (and not 190.8), 190 is selected as it is very close to 190.8.

We will do one more exercise here. If instead of 3” trim size, you select 3.5” trim (6” valve size), what will be the valve opening for the same pressure drop and flow rate?

FCV-01 opening shall be around 60%. Can you figure out, how?

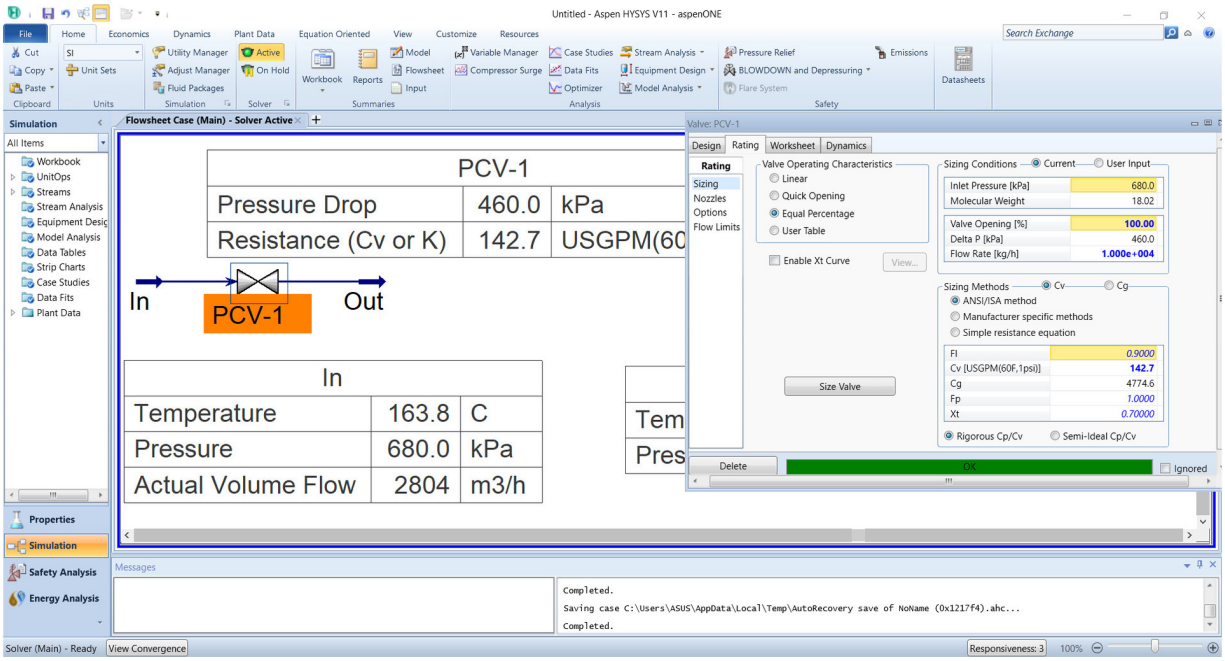
Now let's change the problem statement again. Let the pressure be the same, but instead of sub-cooled liquid, it shall be saturated steam. The flow rate is changed to 10000 kg/h, which is the normal flow rate. It shall be a pressure control valve. Double click on stream "In" and delete temperature value. Provide vapor fraction as 1. It shall automatically calculate T_{sat} at 680 kPa. Change flow rate to 10000 kg/h. Change valve tag to PCV-01.

Figure 7.27: Control valve sizing



Double click on PCV-01. The trim characteristic will be an equal percentage, as it is the pressure control valve. Keep the valve opening at 100%. Select "Sizing Methods" as Cv. Click "Size Valve". The calculated Cv is 142.7.

Figure 7.28: Control valve sizing



Cv of 142.7 is required when the valve is at its normal capacity, but in this case, the actual valve opening shall be about 50% to 60%. Hence, we need to first get the vendor catalog for Equal percentage trim and search for 142.7 Cv value in 50% to 60% valve opening column. (100% valve opening in the above snapshot is correct).

Table 7.5: Cv table for equal percentage trim

Valve Size (in.)	Trim Size (in.)	Cv At Percent Open										
		5	10	20	30	40	50	60	70	80	90	100
0.5	0.50	0.13	0.25	0.39	0.56	0.91	1.4	1.9	2.6	3.7	4.7	5.1
	0.31	0.049	0.10	0.15	0.22	0.34	0.59	1.0	1.2	1.8	2.6	2.9
	0.25B	0.007 7	0.015	0.032	0.053	0.10	0.16	0.25	0.40	0.63	1.0	1.3
0.75	0.72	0.32	0.63	1.0	1.6	2.3	3.2	3.9	5.6	7.6	9.0	9.5
	0.50	0.12	0.24	0.37	0.62	1.0	1.5	1.9	3.0	4.6	5.8	6.4
	0.31	0.046	0.092	0.15	0.20	0.31	0.58	0.87	0.92	1.2	1.8	2.0
	0.25B	0.007 7	0.015	0.032	0.053	0.10	0.16	0.25	0.40	0.63	1.0	1.3
	0.12 A	0.003 0	0.0060	0.018	0.036	0.056	0.088	0.15	0.24	0.37	0.48	0.47
1	0.81	0.53	1.1	1.6	2.1	2.8	3.3	4.9	8.0	11.2	14.1	15.5
	0.62	0.18	0.35	0.61	1.0	1.5	2.3	2.7	4.2	6.5	8.7	10.2
	0.38	0.044	0.088	0.14	0.20	0.355	0.61	0.92	1.1	1.6	2.4	2.9
	0.25B	0.007 7	0.015	0.032	0.053	0.10	0.16	0.25	0.40	0.63	1.0	1.3
	0.12 A	0.013	0.026	0.043	0.063	0.085	0.11	0.16	0.22	0.29	0.46	0.47
1.5	1.25	0.77	1.5	2.3	3.5	4.9	6.6	10	15	21	25	27
	0.81	0.17	0.33	0.69	1.1	1.8	2.6	4.0	6.3	8.9	11	13
	0.38	0.054	0.11	0.18	0.30	0.58	0.86	1.2	1.4	1.9	2.8	3.6
2	1.62	1.3	2.6	3.8	5.9	9.0	11	16	25	34	41	46
	1.00	0.53	1.1	1.6	2.5	3.3	4.7	7.5	11	15	19	21
	0.38	0.075	0.15	0.22	0.34	0.55	0.91	1.1	1.7	2.6	3.2	3.5
3	2.62	2.9	5.8	10	16	22	38	60	76	85	95	104
	1.62	1.2	2.5	3.6	5.7	8.4	11	16	24	36	46	52
4	3.50	6.5	15	20	26	35	57	96	150	154	169	179
	2.25	2.3	4.6	7.4	12	15	24	40	63	83	96	104
6	5.00	6.9	14	20	33	39	125	202	264	302	333	335
	2.62	3.5	6.9	11	16	22	37	65	95	118	122	141

As per the above table, 6” valve with 5” trim shall be adequate for the given application.

Now let’s open Example 1.hsc and change Fl value from 0.9 to 0.6, meaning the control valve has less Liquid pressure recovery factor. This increases valve Cv and makes the valve choked. Hence, whenever you encounter any valve with the choking problem, first check its Fl value and calculate if that is causing the choking. If yes, then replace the valve with anti-cavitation trim (higher Fl value valve).

Figure 7.29: Control valve sizing

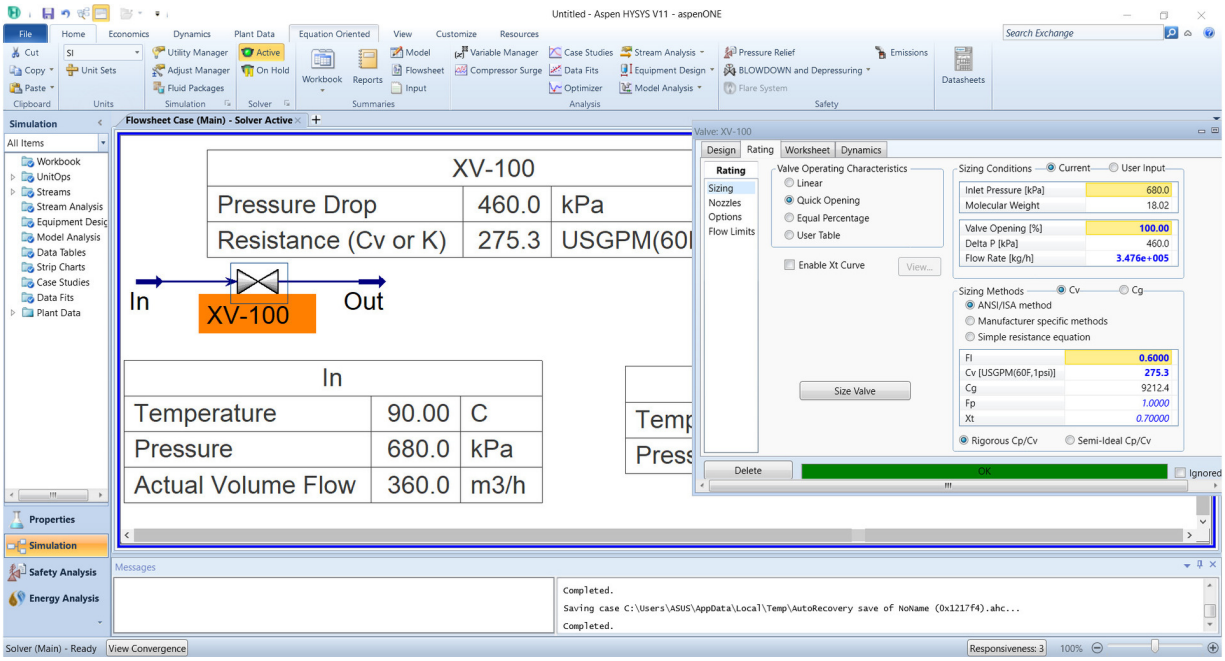
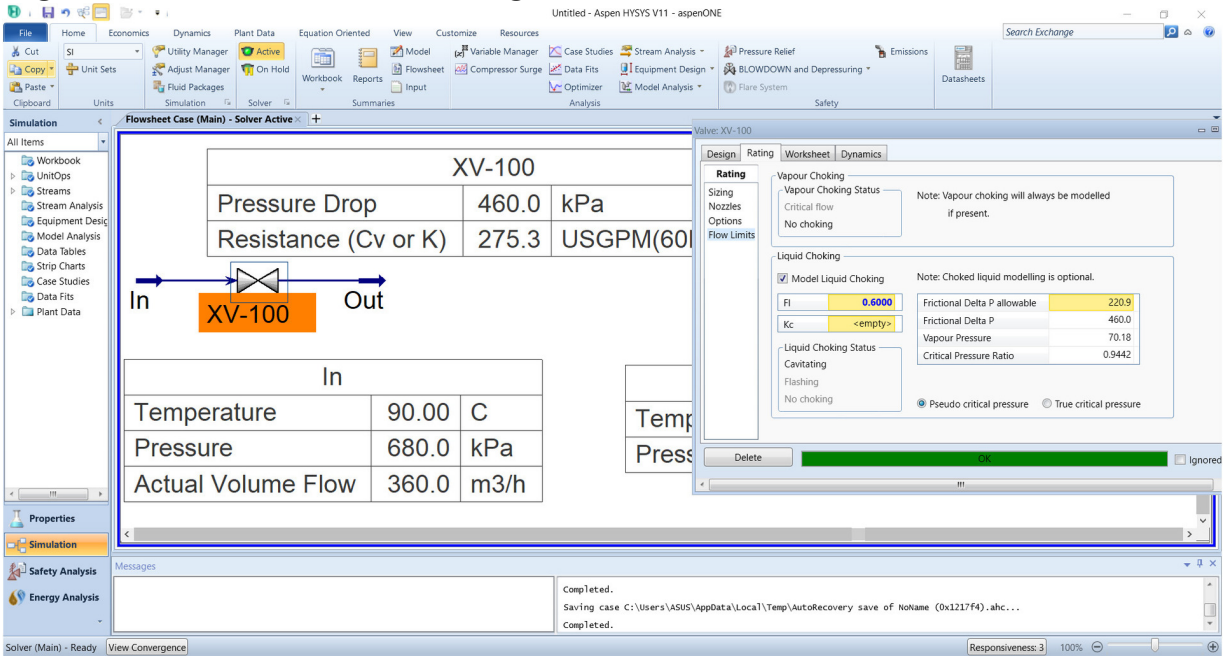

Untitled - Aspen HYSYS V11 - aspenONE
Flowsheet Case (Main) - Solver Active
XV-100
Pressure Drop 460.0 kPa
Resistance (Cv or K) 275.3 USGPM(60
In
XV-100
Out
In
Temperature 90.00 C
Pressure 680.0 kPa
Actual Volume Flow 360.0 m3/h
Valve: XV-100
Design Rating Worksheet Dynamics
Valve Operating Characteristics
Linear
Quick Opening
Equal Percentage
User Table
Enable Xt Curve
Size Valve
Sizing Conditions Current User Input
Inlet Pressure [kPa] 680.0
Molecular Weight 18.02
Valve Opening [%] 100.00
Delta P [kPa] 460.0
Flow Rate [kg/h] 3.476e+005
Sizing Methods Cv Cg
ANSI/ISA method
Manufacturer specific methods
Simple resistance equation
Fl 0.6000
Cv [USGPM(60F,1psi)] 275.3
Cg 9212.4
Fp 1.0000
Xt 0.70000
Rigorous Cp/Cv
Semi-Ideal Cp/Cv
Delete
OK
Ignored
Messages
Completed.
Saving case C:\Users\ASUS\AppData\Local\Temp\AutoRecovery save of NoName (0x1217f4).ahc...
Completed.
Solver (Main) - Ready
View Convergence
Responsiveness: 3
100%

Figure 7.30: Control valve sizing



When the process engineer does not have access to vendor catalog for concept or FEED stage study, then control valve size can be determined from below rule of thumb.

$Cv = 10\ d^2$

Where d is valve size in inch.
For example, a 2" valve Cv will be around 40 and 6" valve Cv will be around 360.

A word of caution, as a process engineer, you may be able to recommend the adequate valve size, but you should not do that. The control valve vendor should be the one responsible to use the input data given by the process engineer and do the calculation and recommend the appropriate valve size. The knowledge that you have gained here, must be used to verify/review vendor design and make sure your valve size and vendor's valve size is the same. The vendor provides a process guarantee for the control valve, not you.

ANSI/ISA 75.01.01 should be used to prepare control valve sizing spreadsheets for compressible and incompressible fluids. All the formula, methodology, and flow charts are given in the same standard.

Exercises

a. Ms. Vasantha is learning about cavitation in the control valve. Please make a presentation file that explains the same in detail.
b. Mr. Koteswararao is not sure how to find out the Cv of a valve. Have a brainstorming session in your class/office and try to find the answer.
c. Mr. Nitish is trying to select appropriate PCV size for natural gas at 50 ^{o} C, 5000 kg/h (design flow), 6.5 barg upstream conditions, and with 0.5 bar pressure drop across the valve. Can you help him?
d. Mr. Abizer needs to make a presentation about different trim characteristics for the control valve and compare it with the ones given by valve vendors. Please help him.
e. Mr. Rahul is trying to select appropriate LCV size for toluene at 20 ^{o} C, 55 m^3 /h (normal flow), 5 barg upstream conditions, and with 1 bar pressure drop across the valve. Can you help him?
f. Mr. Uday is preparing a presentation about self-acting regulators. Please help him to prepare such a presentation file.

CHAPTER 8: ORIFICE SIZING USING ASPEN HYSYS

An Orifice meter is a type of flow meter used to measure the rate of flow of fluid using differential pressure measurement principle. An orifice plate is nothing but a thin hole with a hole that is placed in a pipe in which fluid flows. As fluid flows through the orifice, it causes pressure drop due to reduced flow area. More flow rate means more pressure drop. Hence for a constant flow area of the orifice, as pressure drop is measured, then flow rate can be calculated.

Figure 8.1: Orifice

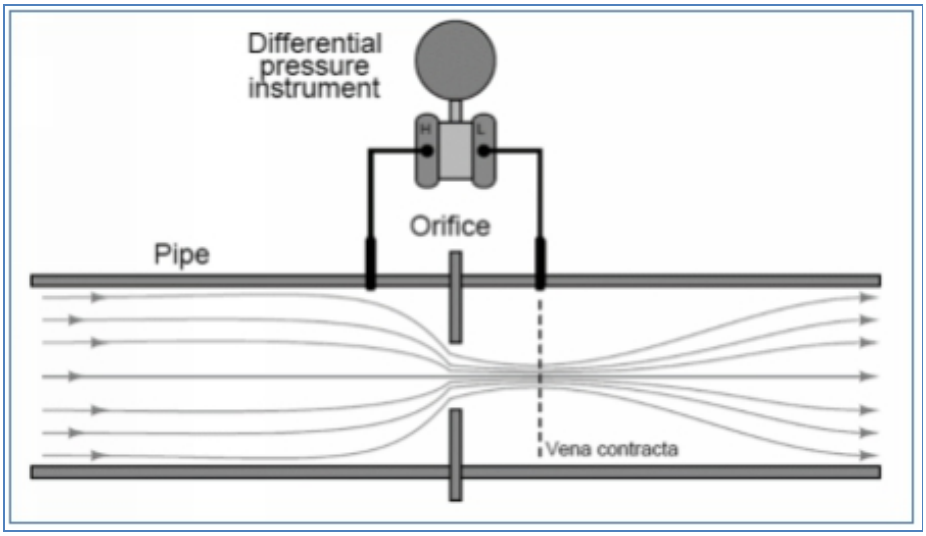
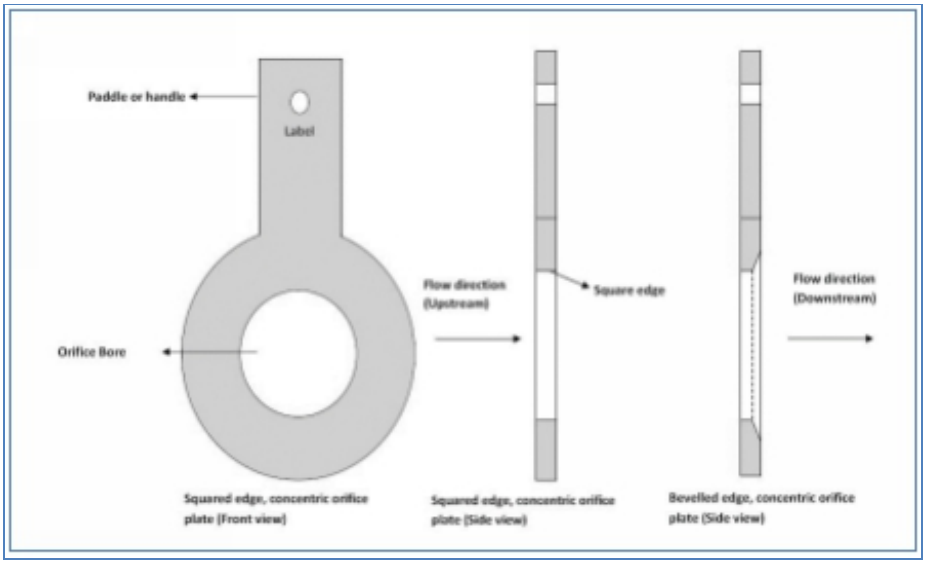


Figure 8.2: Orifice



ISO 5167-2 standard is used for calculation of orifice sizing. It applies only to a flow that remains subsonic throughout the measuring section and where the fluid can be considered as a single phase. It does not apply to the measurement of pulsating flow. It does not cover the use of orifice plates in pipe sizes less than 50 mm or more than 1 000 mm, or for pipe Reynolds numbers below 5000. The diameter d shall in all cases be greater than or equal to 12,5 mm. The diameter ratio, d/D, shall be always greater than or equal to 0,10 and less than or equal to 0,75.
Flow orifice usually has a pressure drop of up to 0.5 bar. Restriction orifice has a significant pressure drop.

Types of Restriction Orifice Plates

- **Single-stage restriction orifice**

A single-stage restriction orifice is usually a plate or a block with a bore (orifice) sized to the intended permanent loss of pressure. It is installed between the pipe flanges. Usually, it is not a thin orifice plate; it is a thick orifice plate.

- **Single-stage multi-hole restriction orifice**

A single-stage multi-hole restriction orifice plate is used to abate the noise generated by the device due to high velocity through the bore which offers restriction to the incoming fluid.
The flow at the inlet is now channeled into several streams through the multiple holes and this reduces the noise which would be otherwise will be above the acceptable limit if a single hole device is used.

- **Multi-stage restriction orifice plate assembly**

These devices are used where the pressure reduction ratio is very high and cannot be achieved by a single-

stage orifice plate. Thus, a multistage device essentially consists of several single-stage devices built in a single spool.

Like a single-stage device, it can be of single hole multistage design or multi-hole multi-stage design or combination of both.

Simulation

Let's solve an example using Aspen HYSYS.

Water flows at 60000 kg/h across a flow orifice with an assigned pressure drop value of 0.5 bar. Water upstream pressure is 5 barg. Pipe ID to be considered as 100 mm. Calculate orifice area, orifice diameter, and beta ratio.

Select Water as a component and IAPWS as a Fluid package.

Figure 8.3: Component list

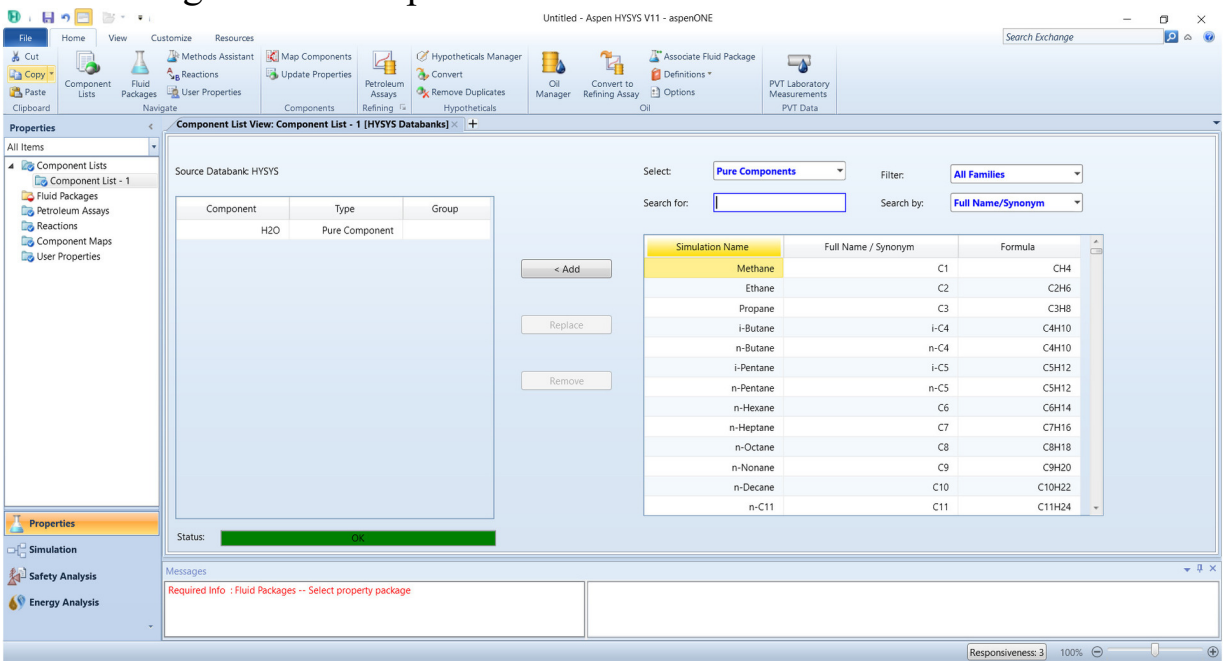
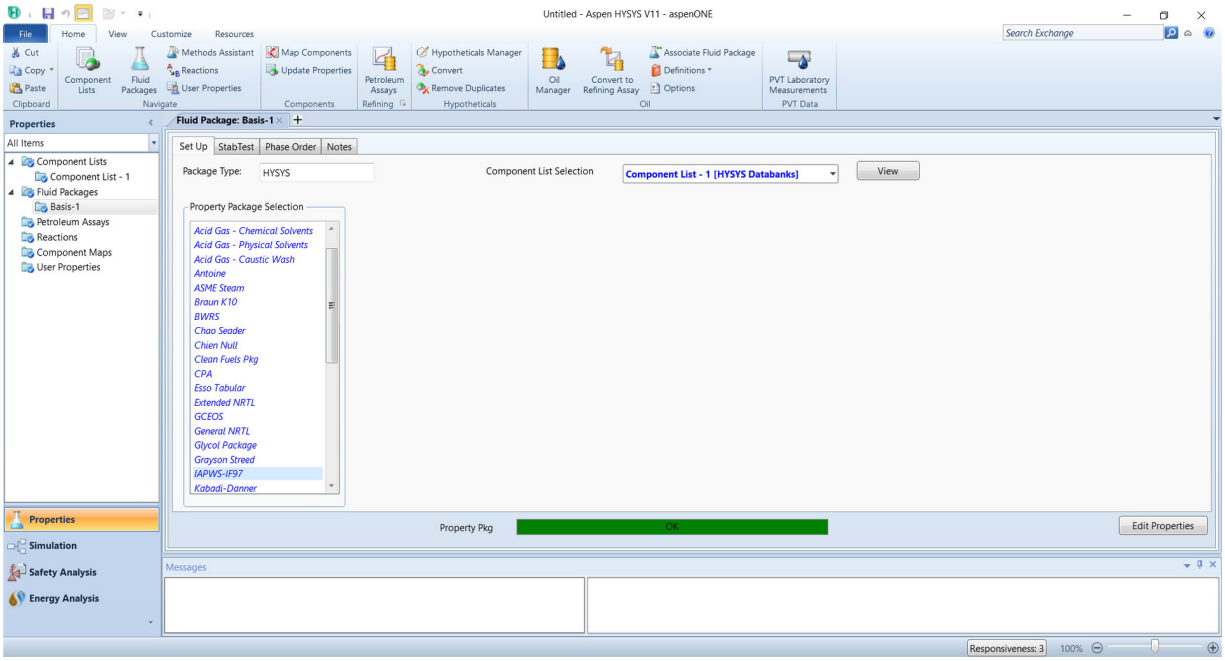


Figure 8.4: Fluid package


Untitled - Aspen HYSYS V11 - aspenONE
File
Home
View
Customize
Resources
Search Exchange
Cut
Copy
Paste
Clipboard
Component Lists
Fluid Packages
Methods Assistant
Reactions
User Properties
Navigate
Map Components
Update Properties
Components
Petroleum Assays
Refining
Hypotheticals Manager
Convert
Remove Duplicates
Hypotheticals
Oil Manager
Convert to Refining Assay
Associate Fluid Package
Definitions
Options
Oil
PVT Laboratory Measurements
PVT Data
Properties
All Items
Component Lists
Component List - 1
Fluid Packages
Basis-1
Petroleum Assays
Reactions
Component Maps
User Properties
Fluid Package: Basis-1
Set Up
StabTest
Phase Order
Notes
Package Type:
HYSYS
Component List Selection
Component List - 1 [HYSYS Databanks]
View
Property Package Selection
Acid Gas - Chemical Solvents
Acid Gas - Physical Solvents
Acid Gas - Caustic Wash
Antoine
ASME Steam
Braun K10
BWRS
Chao Seader
Chien Null
Clean Fuels Pkg
CPA
Esso Tabular
Extended NRTL
GCEOS
General NRTL
Glycol Package
Grayson Streed
IAPWS-IF97
Kabadi-Danner
Property Pkg
OK
Edit Properties
Properties
Simulation
Safety Analysis
Energy Analysis
Messages
Responsiveness: 3
100%

Insert a Material Stream from Model Palette and provide temperature, pressure, flow rate, and composition as shown below.

Figure 8.5: Material stream

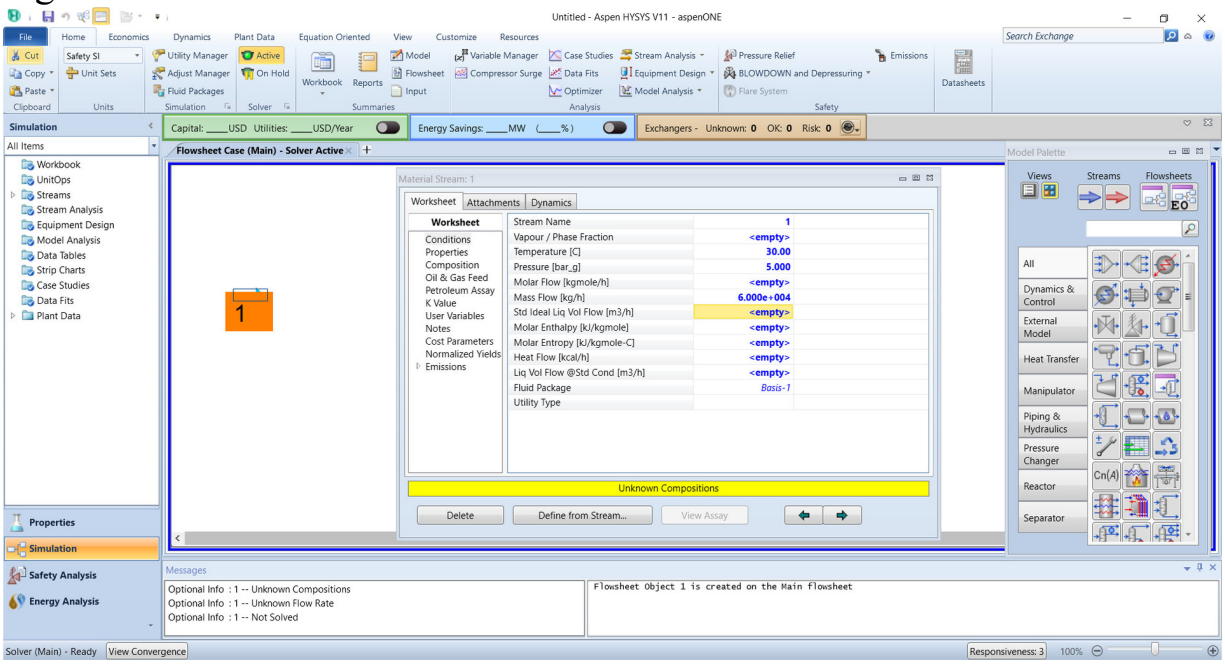
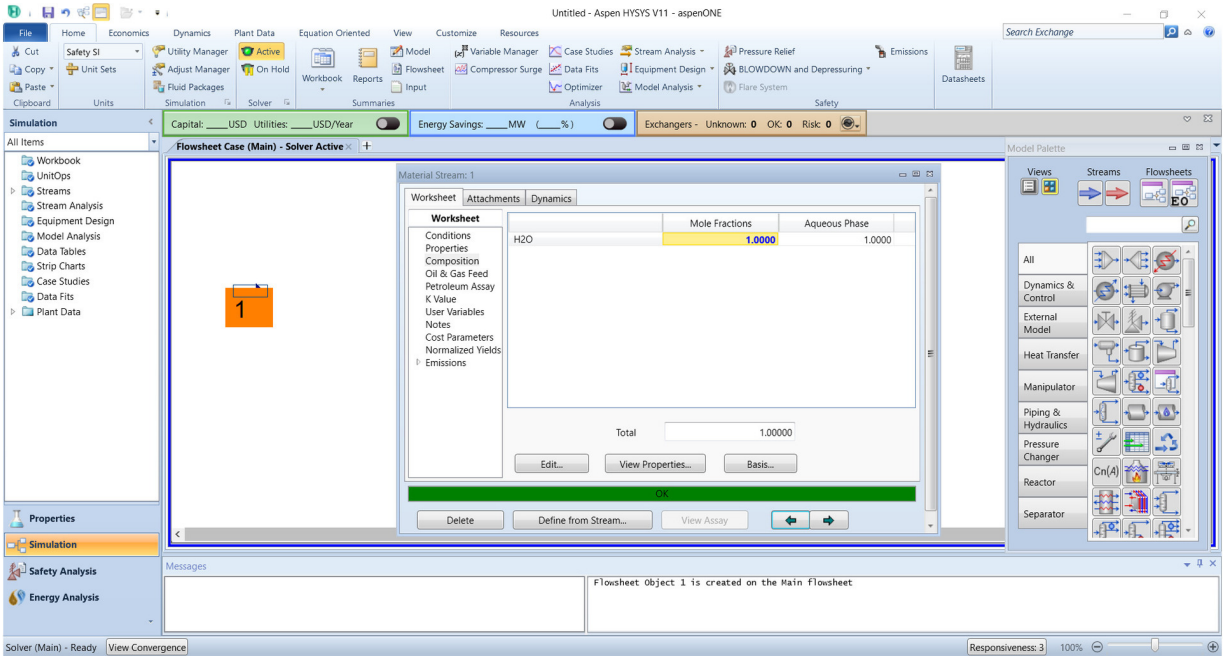
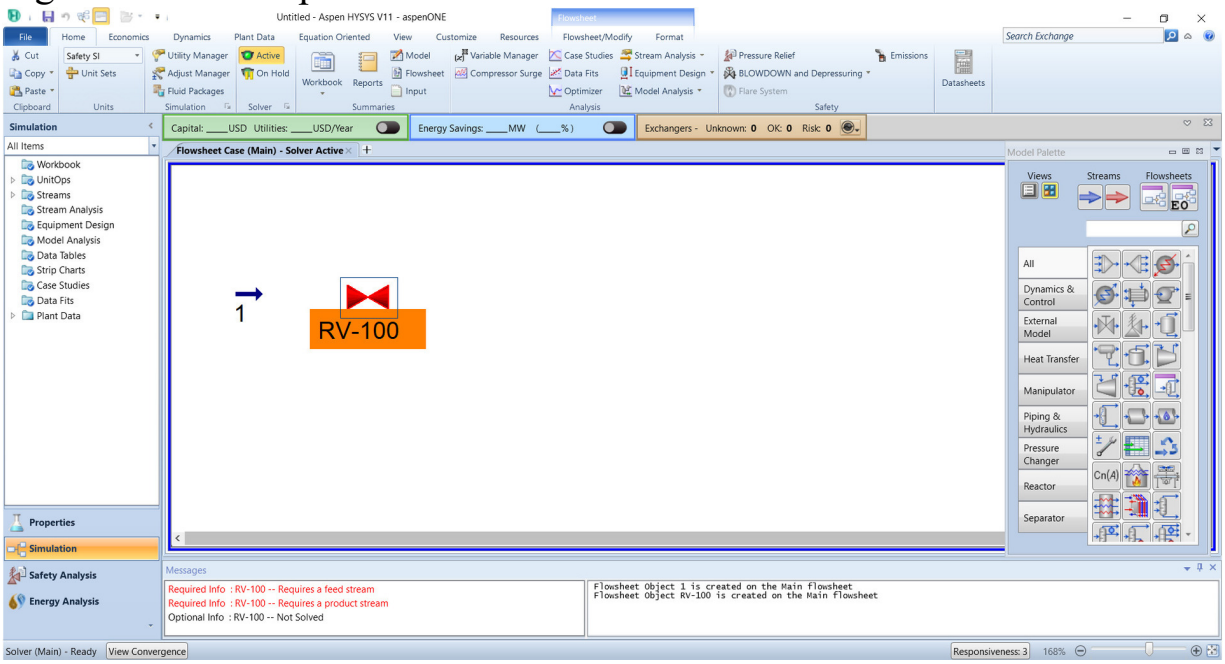


Figure 8.6: Material stream



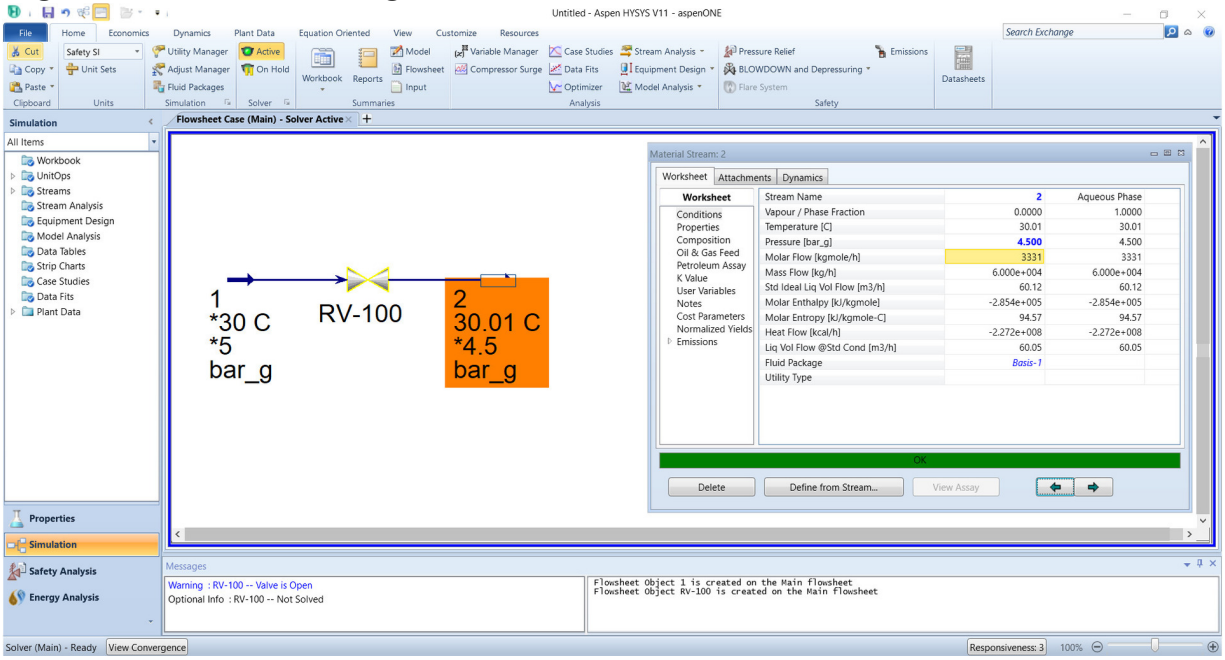
Insert “Relief Valve” from the Model Palette and connect the streams. “Relief valve” can be used to simulate orifice.

Figure 8.7: Model palette



Provide outlet stream pressure as 4.5 barg.

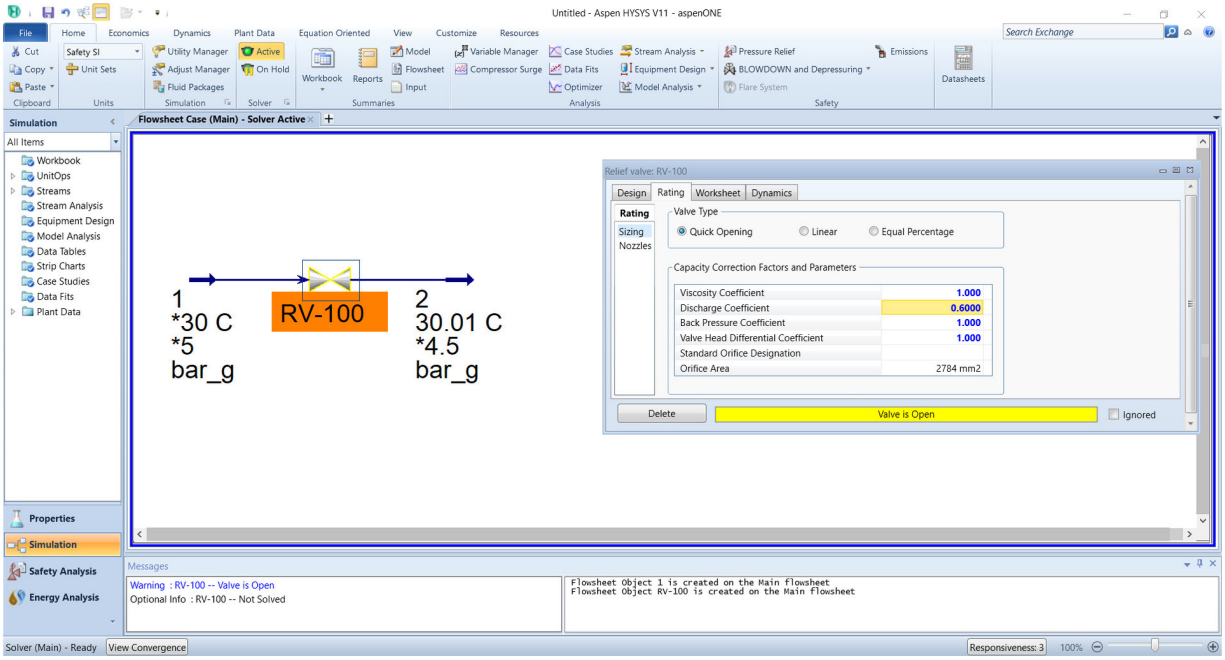
Figure 8.8: Orifice sizing



Double-click on Orifice (RV-100) and provide a discharge coefficient (0.6) in the Rating tab.

The orifice area calculated is 2784 mm^2 . Hence 60 mm orifice diameter (d) is calculated. Pipe inside diameter (D) is 100 mm. Beta ratio (d/D) is 0.6, which is within the allowable range of 0.1 to 0.75 as per ISO 5167-2. Hence calculated orifice size is adequate.

Figure 8.9: Orifice sizing



Exercises

a. Mr. Rajesh is trying to select appropriate orifice size for natural gas at 50 ^{o} C, 5000 kg/h, 6.5 barg upstream conditions, and with 0.5 bar pressure drop across orifice. Can you help him?
b. Ms. Neeta is trying to select appropriate orifice size for natural gas at 50 ^{o} C, 5000 kg/h, 6.5 barg upstream conditions, and with 6 bar pressure drop across orifice. Can you help her?
c. Mr. Swapnil is trying to select appropriate orifice size for Benzene at 10 ^{o} C, 35 m^3 /h, 3.5 barg upstream conditions, and with 0.5 bar pressure drop across orifice. Can you help him?

CHAPTER 9: SEPARATOR SIZING USING ASPEN HYSYS

Separator is a pressure vessel used for separating well fluids from oil & gas wells into gaseous and liquid components. A separator for petroleum production is a large vessel designed to separate production fluids into the components of oil, gas, and water.

Types of Separators

Flash Vessel

A vessel used to separate the gas evolved from liquid flashed from a higher pressure to a lower pressure.

Gas-Liquid-Solid Separator (Filter Separator)

A filter separator usually has two compartments. The first compartment contains filter-coalescing elements. The liquid particles coalesce into large droplets and when the droplets reach sufficient size, the gas flow causes them to flow out of the filter elements into the center core. The particles are then carried into the second compartment of the vessel (containing a vane-type or knitted wire mesh mist extractor) where the larger droplets are removed.

Knock out drum (Gas-Liquid Separator)

A vessel designed to handle streams with high gas to liquid ratios (GLR). The liquid is generally entrained as mist in the gas or is free-flowing along the pipe wall. These vessels usually have a small liquid collection section.

Two-Phase Separator (Gas-Liquid Separator)

A vessel that separates the well fluid into gas and liquid. A two-phase separator can be horizontal, vertical, or spherical. The liquid leaves the vessel at the bottom through a level control or dump valve. The gas leaves the vessel at the top passing through a mist extractor to remove the entrained liquid droplets in the gas.

Figure 9.1: Vertical separator

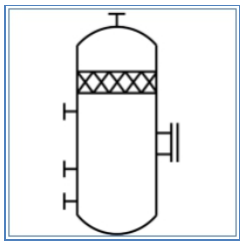
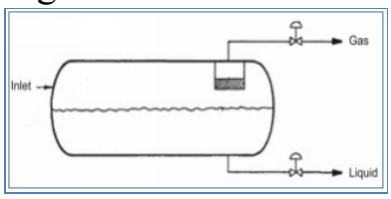


Figure 9.2: Horizontal separator (2 phase)



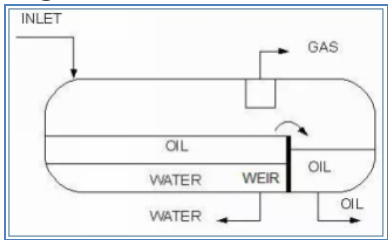
Liquid-Liquid Separator

Two immiscible liquid phases can be separated using the same principles as gas-liquid separators except that they must be designed for much lower feed velocities because of the difference in density between two liquids is less and the separation becomes more difficult.

- **Gas-Liquid (Oil)-Liquid (Water) Separator (Three Phase Separator)**

A vessel that separates the well fluids into gas and two types of liquid oil & water. A three-phase separator can be horizontal, vertical, or spherical. This type of separator is commonly called a free water knock because its main purpose is to remove free water that can cause problems such as corrosion and formation of hydrates.

Figure 9.3: Horizontal separator (3 phase)



Classification by Application

Oil and gas separators are further classified according to the application as below:

Test Separator

A test separator is used to separate and to meter the well fluids. The test separator can be referred to as a well tester. They can be permanently installed or portable. Test separators can be equipped with various types of measuring devices for measuring the Gas oil ratio (GOR), Water cut, Basic sediments & water (BS & W).

Production Separator

Production separator is used to separate the produced well fluid from a well on continuous basis.

Separator Basic Design Criteria

Following three main factors are to be considered in separator sizing:

- Vapor capacity:

The vapor capacity will determine the cross-sectional area necessary for gravitational forces to remove the liquid from the vapor. More vapor load means bigger diameter separator.

- Liquid capacity:

The liquid capacity is set by determining the volume required to provide adequate residence time to "de-gas" the liquid or allow immiscible liquid phases to separate.

- Operability:

Operability issues include the separator's ability to deal with solids if present, unsteady flow, turndown, etc. The optimal design will usually result in an aspect ratio that satisfies these requirements in a vessel of reasonable cost.

One has to follow an iterative approach to the calculations considering the above factors so that the best design is obtained.

As per GPSA, typical vertical H/D ratios are normally in the 2 to 4 range and that of horizontal L/D ratios are normally in the 2.5 to 5 range.

Table 9.1: Function & Separation Device Combinations

Sr. No.	Function	Method of Separation	Internal Devices and Separation Aids
1	Separation of Oil from gas	Gravity	Mist extractors, impingement type, change of direction type, cyclone filter cartridges, and washing.
2	Separation of gas from Oil	Gravity	Cyclone, plate packs, matrix packs, agitation, and heat.
3	Separation of water from Oil	Gravity	Cyclone, plate packs, matrix packs, agitation, and heat.
4	Separation of Oil from water	Gravity	Cyclone, plate packs, matrix packs, agitation, and flotation.
5	Separation of Solids from Liquid	Gravity	Cyclone, agitation, heat, and flotation.

Table 9.2: Separator Orientation

Sr. No.	Operation	Horizontal	Vertical	Discussion
1	Separation	Better	--	In a horizontal vessel, the drops or bubbles do not have to settle or rise through a counter-current flow.
2	Solids	--	Better	Solids are more easily removed from the bottom of the vertical vessel
3	Foaming	Better	--	Horizontal vessel provides more surface area for bubbles to escape
4	Surges	--	Better	A change in liquid level does not affect the gas capacity of the vessel

Mist Extraction Equipment

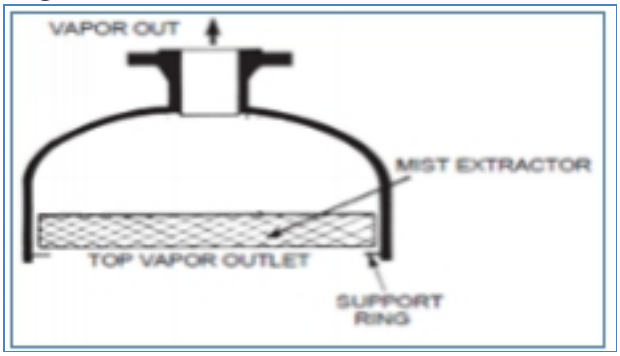
Mist extractors are used to separate the small liquid droplets from the gas phase that were not removed by the inlet device or gas settling section of the separator and the droplets are less than 150-500 microns in size. It is not economic to separate these small droplets by gravity settling.

Wire mesh

Wire-mesh mist extractors, or pads, are made by knitting wire, metal or plastic, into tightly packed layers which are then crimped and stacked to achieve the required pad thickness.

Most mesh pads are 150 mm thick with 144 – 192 kg/m^3 bulk density. Minimum recommended pad thickness is 100 mm. Typical mesh pad installation is shown below.

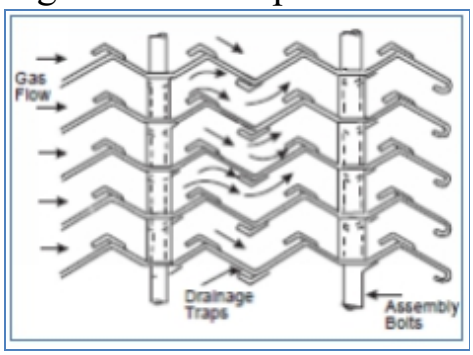
Figure 9.4: Mist extractor



Vane pack

Vane or chevron-type mist extractors (vane-pack) use closely spaced blades arranged to provide sinusoidal or zigzag flow paths. The changes in gas flow direction combined with the inertia of the entrained liquid droplets, cause impingement of the droplets onto the plate surface, followed by coalescence and drainage of the liquid to the liquid collection section of the separator.

Figure 9.5: Vane pack



Cyclonic Separator

Cyclonic separator removes particulates from an air, gas or liquid stream, without the use of filters, due to vortex separation. Gravity and rotational effects are used to separate mixtures of solids and fluids. The method is also used to separate fine droplets of liquid from a gaseous stream.

Souders-Brown Equation

The Souders–Brown equation is used to calculate the maximum allowable vapor velocity in vapor-liquid separation vessels (variously called flash drums, knockout drums, knockout pots, compressor suction drums, and compressor inlet drums). The equation is given below;

$$V = (k)\sqrt{\frac{\rho_L - \rho_V}{\rho_V}}$$

Where:

- V = maximum allowable vapor velocity, m/s
- ρ_L = liquid density, kg/m³
- ρ_V = vapor density, kg/m³
- K = mesh capacity factor

The separator has a vapor outlet at the top, liquid outlet at the bottom, and feed inlet at about the half-full level. At the vapor outlet, a mesh pad is provided. The liquid outlet line may have a liquid level control valve.

Recommended values of k

The GPSA Engineering Data Book recommends the following k values for vertical drums with horizontal mesh pads (at the denoted operating pressures):

- At a gauge pressure of 0 bar: 0.107 m/s
- At a gauge pressure of 7 bar: 0.107 m/s
- At a gauge pressure of 21 bar: 0.101 m/s
- At a gauge pressure of 42 bar: 0.092 m/s
- At a gauge pressure of 63 bar: 0.083 m/s
- At a gauge pressure of 105 bar: 0.065 m/s

Notes:

- k = 0.107 at a gauge pressure of 7 bar. Subtract 0.003 for every 7 bar above a gauge pressure of 7 bar.
- For glycol or amine solutions, multiply above k values by 0.6 – 0.8.
- Typically use one-half of the above k values for approximate sizing of vertical separators without mesh pads
- For compressor suction scrubbers and expander inlet separators, multiply k by 0.7 – 0.8.

Some guidelines for separator design are provided below.

Table 9.3: Momentum & Velocity criteria for nozzles (Source: DEP 31.22.05.12 - Gen.- 2008)

Inlet Nozzle	**Momentum or Velocity Limit**
No Internal	Max. Momentum, $\varrho_m V^2_m \leq 1400$ Pa
Half-open pipe	Max. Momentum, $\varrho_m V^2_m \leq 2100$ Pa
Schoepentoeter used as inlet device	Max. Momentum, $\varrho_m V^2_m \leq 8000$ Pa
Gas outlet Nozzle	Max. Momentum, $\varrho_g V^2_{g,out} \leq 4500$ Pa
Liquid outlet Nozzle	Maximum velocity, 1 m/s

Table 9.4: Selection guideline for separator types

System Characteristics	Type of Separator
Large vapor, less liquid Load (by volume)	Vertical
Large liquid, less vapor Load (by volume)	Horizontal
Large vapor, large liquid Load (by volume)	Horizontal
Liquid-liquid separation	Horizontal
Liquid-solid separation	Vertical

Table 9.5: Level setting in the separator

Level type	Level setting
Level Alarm High High (LAHH)	30 – 60 seconds or 200 mm whichever is greater
Level Alarm High (LAH)	30 – 60 seconds or 200 mm whichever is greater
Normal Alarm Level (NAL)	60% of horizontal separator
Level Alarm Low (LAL)	30 – 60 seconds or 200 mm whichever is greater
Level Alarm Low Low (LALL)	30 – 60 seconds or 200 mm whichever is greater Should be at least 200 mm above the vessel bottom or maximum interface level

Table 9.6: Typical K factors for the sizing of wire mesh demisters (Source: IPS-E-PR-880, 1997)

Separator type	K factor (m/s)
Horizontal (with vertical pad)	0.122 to 0.152
Spherical	0.061 to 0.107
Vertical or horizontal (with horizontal pad)	0.055 to 0.107
At atmospheric pressure	0.107
At 2100 kPa	0.101
At 4100 kPa	0.091
At 6200 kPa	0.082
At 10300 kPa	0.064
Wet steam	0.076
Most vapors under vacuum	0.061
Salt and caustic evaporators	0.046

Table 9.7: Standard separator sizes as per API

D [in] x H or L [ft]		
12¾ in x 5 ft		36 in x 7½ ft
12¾ in x 7½ ft	36 in x 10 ft	
12¾ in x 10 ft	36 in x 15 ft	
16 in x 5 ft	42 in x 7½ ft	

D [in] x H or L [ft]		
16 in x 7½ ft	42 in x 10 ft	
16 in x 10 ft	42 in x 15 ft	
20 in x 5 ft	48 in x 7½ ft	

D [in] x H or L [ft]		
20 in x 7½ ft	48 in x 10 ft	
20 in x 10 ft	48 in x 15 ft	
24 in x 5 ft	54 in x 7½ ft	

D [in] x H or L [ft]		
24 in x 7½ ft	54 in x 10 ft	
24 in x 10 ft	54 in x 15 ft	
30 in x 5 ft	60 in x 7½ ft	

D [in] x H or L [ft]		
30 in x 7½ ft	60 in x 10 ft	
30 in x 10 ft	60 in x 15 ft	
36 in x 5 ft		

Simulation

Let's solve a separator sizing problem using Aspen HYSYS.

Inlet to the separator is a mixture of Methane and n-Butane. Separator operating temperature and pressure are 80 ° F and 814.7 psia respectively. The flow rate is 3221 lbmol/h. Methane 81.32% (mole) and 18.68% (mole) n-Butane. The fluid package is Peng Robinson.

Figure 9.6: Component list

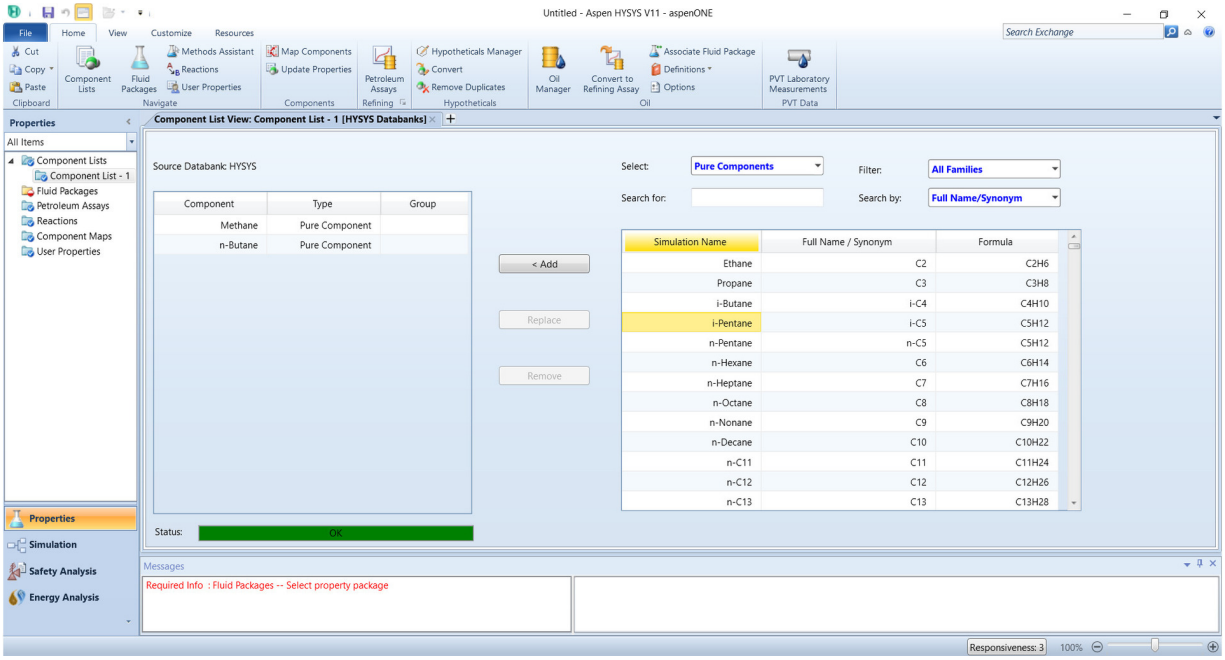
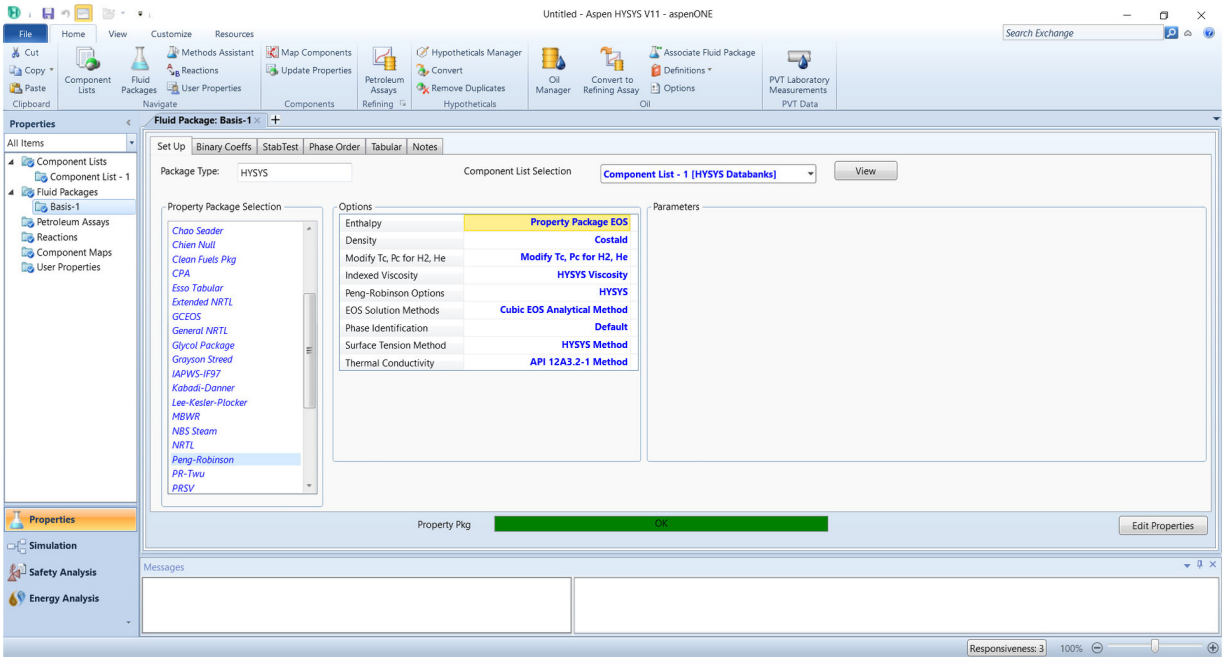
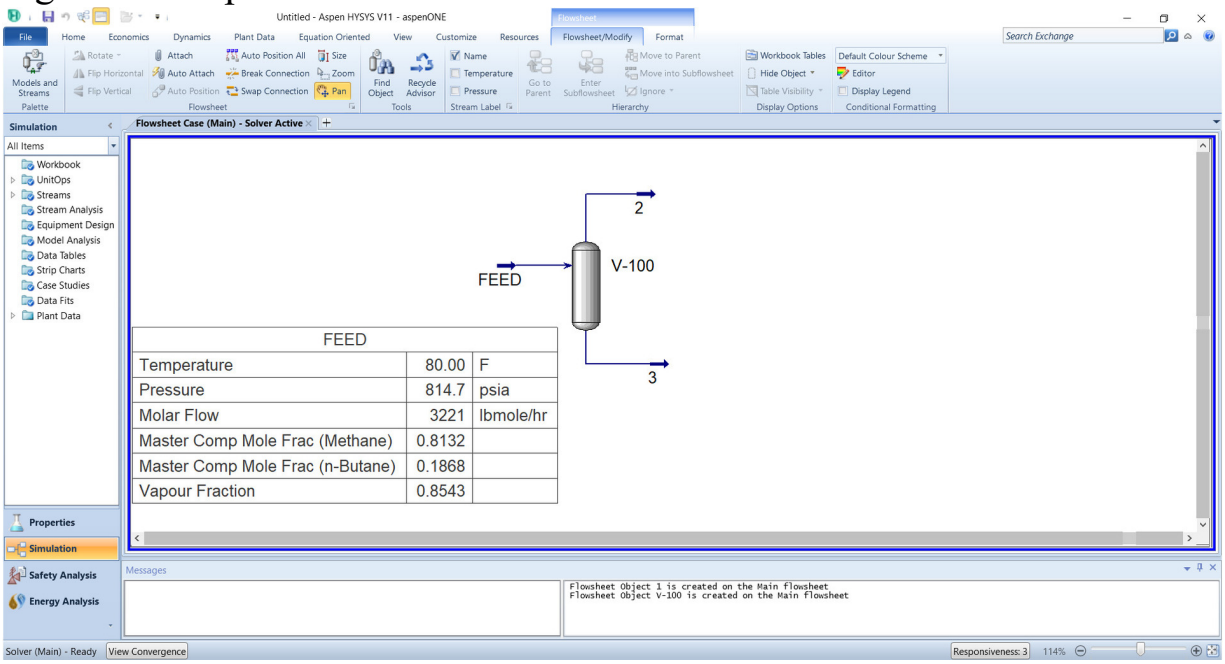


Figure 9.7: Fluid package


Untitled - Aspen HYSYS V11 - aspenONE
File
Home
View
Customize
Resources
Search Exchange
Cut
Copy
Paste
Clipboard
Component Lists
Fluid Packages
Methods Assistant
Reactions
User Properties
Navigate
Map Components
Update Properties
Components
Petroleum Assays
Refining
Hypotheticals Manager
Convert
Remove Duplicates
Hypotheticals
Oil Manager
Convert to Refining Assay
Associate Fluid Package
Definitions
Options
Oil
PVT Laboratory Measurements
PVT Data
Properties
All Items
Component Lists
Component List - 1
Fluid Packages
Basis-1
Petroleum Assays
Reactions
Component Maps
User Properties
Fluid Package: Basis-1
Set Up
Binary Coeffs
StabTest
Phase Order
Tabular
Notes
Package Type:
HYSYS
Component List Selection
Component List - 1 [HYSYS Databanks]
View
Property Package Selection
Chao Seader
Chien Null
Clean Fuels Pkg
CPA
Esso Tabular
Extended NRTL
GCEOS
General NRTL
Glycol Package
Grayson Streed
IAPWS-IF97
Kabadi-Danner
Lee-Kesler-Plocker
MBWR
NBS Steam
NRTL
Peng-Robinson
PR-Twu
PRSV
Options
Enthalpy
Property Package EOS
Density
Costald
Modify Tc, Pc for H2, He
Modify Tc, Pc for H2, He
Indexed Viscosity
HYSYS Viscosity
Peng-Robinson Options
HYSYS
EOS Solution Methods
Cubic EOS Analytical Method
Phase Identification
Default
Surface Tension Method
HYSYS Method
Thermal Conductivity
API 12A3.2-1 Method
Parameters
Property Pkg
OK
Edit Properties
Properties
Simulation
Safety Analysis
Energy Analysis
Messages
Responsiveness: 3
100%

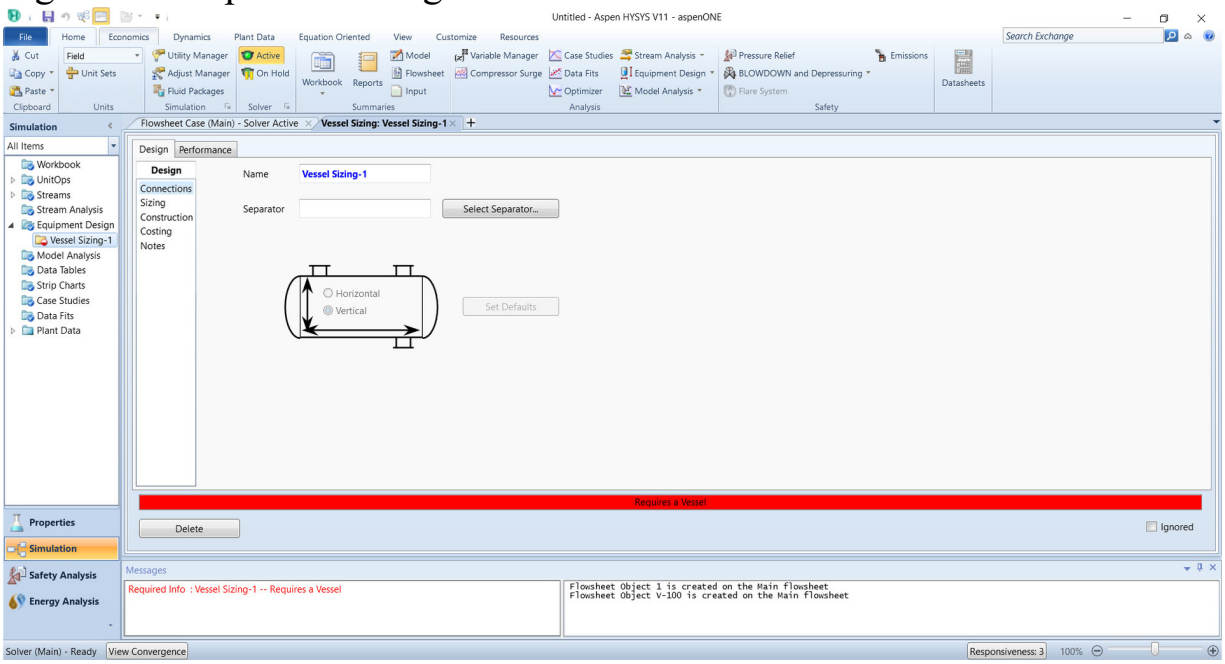
Select the Separator from the model palette and connect the material streams.

Figure 9.8: Separator



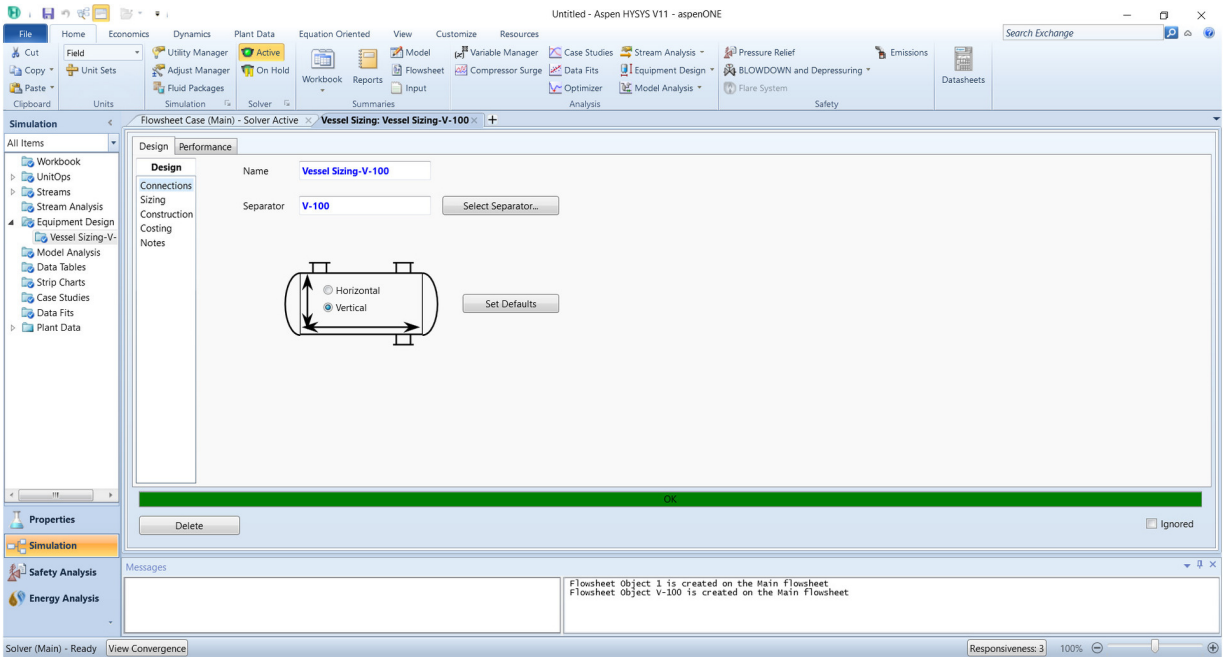
Select “Home\Analysis\Equipment Design” drop-down box and choose “Vessel Sizing”.

Figure 9.9: Separator sizing



Click on “Select Separator” as V-100 and select “Vertical” separator. As per the best engineering practice, more vapor feed should be sent to a vertical separator.

Figure 9.10: Separator sizing



Select the "Sizing" tab and add "Diameter" as an "Additional Specifications". For vertical separators, the H/D ratio is 2 to 4. Liquid residence time will depend on the ° API of the liquid. ° API can be calculated using the specific gravity of liquid using the below formula.

$$\text{API gravity} = \frac{141.5}{\text{SG}} - 131.5$$

Typical retention time v/s Oil gravity is given in the below table.

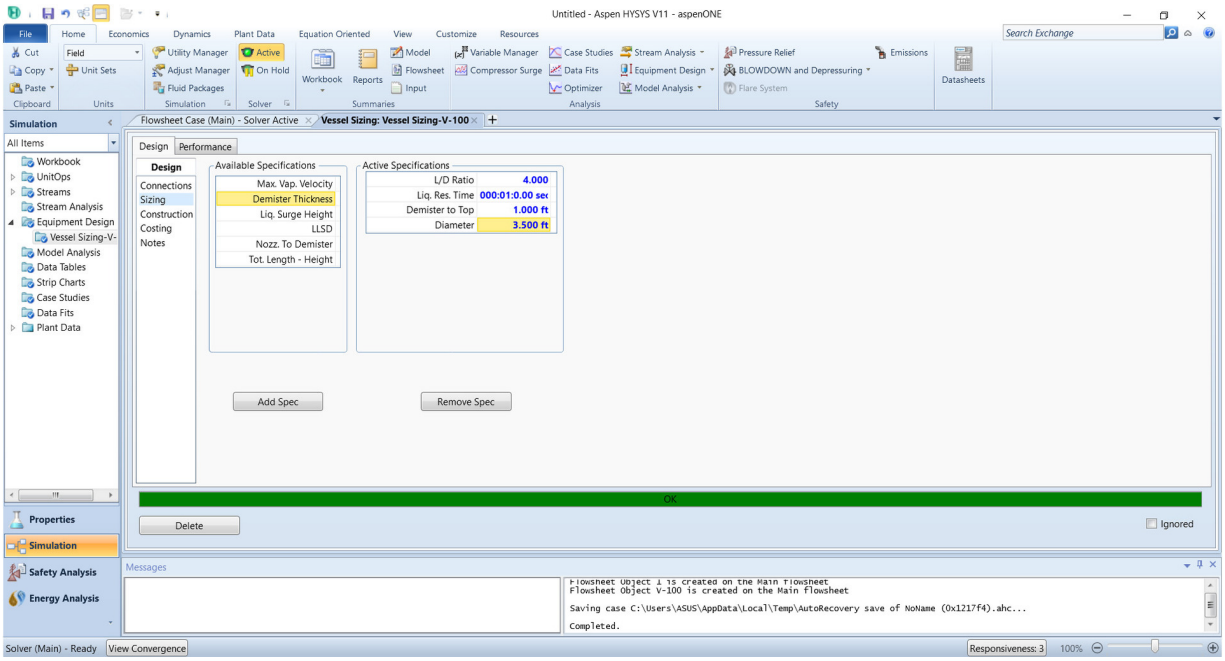
Table 9.8: Retention time guideline

Oil Gravities	**Retention time, minutes (Typical)**
Above 35 ° API	1
20 to 30 ° API	1 to 2
10 to 20 ° API	2 to 4

As per the above table, retention time in this example will be 1 minute as the specific gravity of a liquid is 0.49 as per HYSYS and calculated ° API shall be 157.

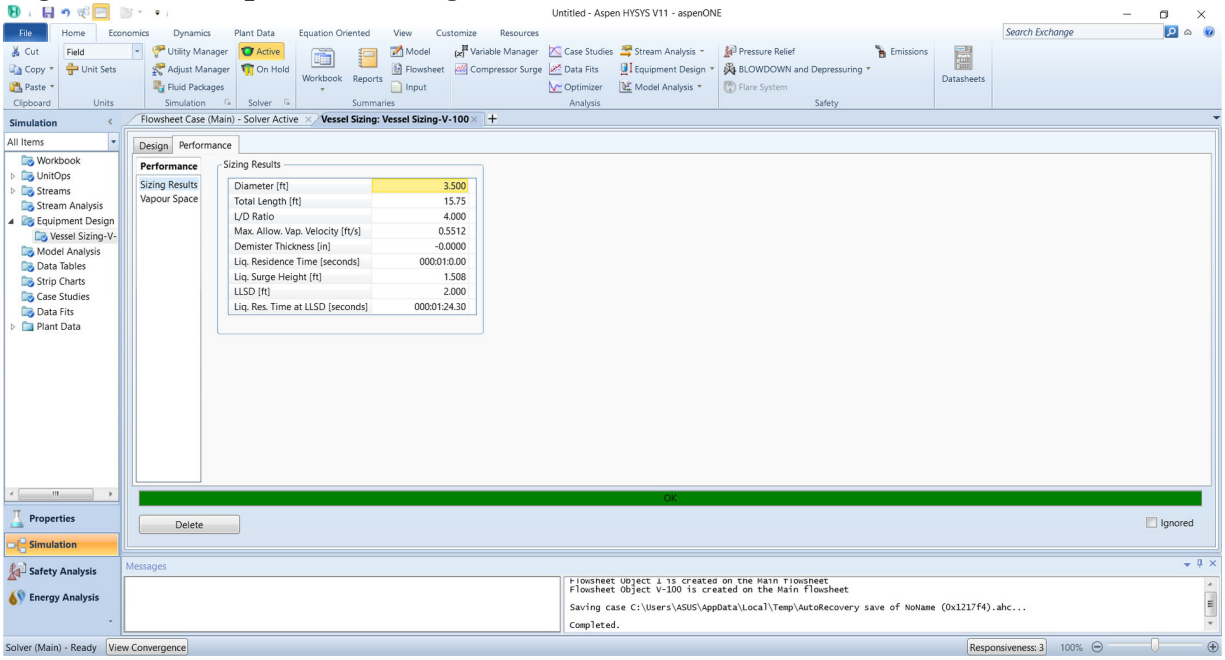
We will keep the separator diameter as 3.5 ft.

Figure 9.11: Separator sizing


Untitled - Aspen HYSYS V11 - aspenONE
File
Home
Economics
Dynamics
Plant Data
Equation Oriented
View
Customize
Resources
Search Exchange
Cut
Copy
Paste
Clipboard
Field
Unit Sets
Units
Utility Manager
Adjust Manager
Fluid Packages
Simulation
Active
On Hold
Solver
Workbook
Reports
Model
Flowsheet
Input
Summaries
Variable Manager
Compressor Surge
Case Studies
Data Fits
Optimizer
Stream Analysis
Equipment Design
Model Analysis
Analysis
Pressure Relief
BLOWDOWN and Depressuring
Flare System
Emissions
Safety
Datasheets
Simulation
All Items
Workbook
UnitOps
Streams
Stream Analysis
Equipment Design
Vessel Sizing-V-
Model Analysis
Data Tables
Strip Charts
Case Studies
Data Fits
Plant Data
Properties
Simulation
Safety Analysis
Energy Analysis
Flowsheet Case (Main) - Solver Active
Vessel Sizing: Vessel Sizing-V-100
Design
Performance
Design
Connections
Sizing
Construction
Costing
Notes
Available Specifications
Max. Vap. Velocity
Demister Thickness
Liq. Surge Height
LLSD
Nozz. To Demister
Tot. Length - Height
Active Specifications
L/D Ratio 4.000
Liq. Res. Time 000:01:0.00 sec
Demister to Top 1.000 ft
Diameter 3.500 ft
Add Spec
Remove Spec
OK
Delete
Ignored
Messages
Flowsheet Object 1 is created on the Main Flowsheet
Flowsheet Object V-100 is created on the Main Flowsheet
Saving case C:\Users\ASUS\AppData\Local\Temp\AutoRecovery save of NoName (0x1217f4).ahc...
Completed.
Solver (Main) - Ready
View Convergence
Responsiveness: 3
100%

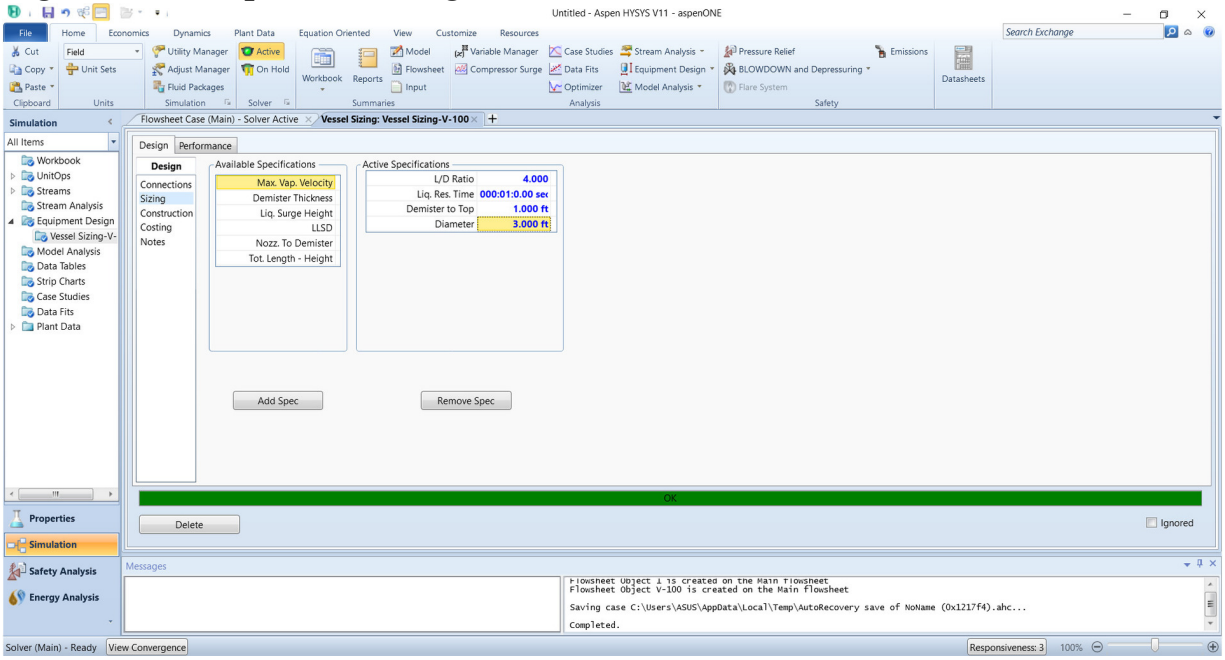
Accordingly, results are displayed in the “Performance” tab. Total height with dished end is 15.75 ft. Liquid level is 1.5 ft, meaning % liquid level will be 1.5/15.75x100 = 10%.

Figure 9.12: Separator sizing



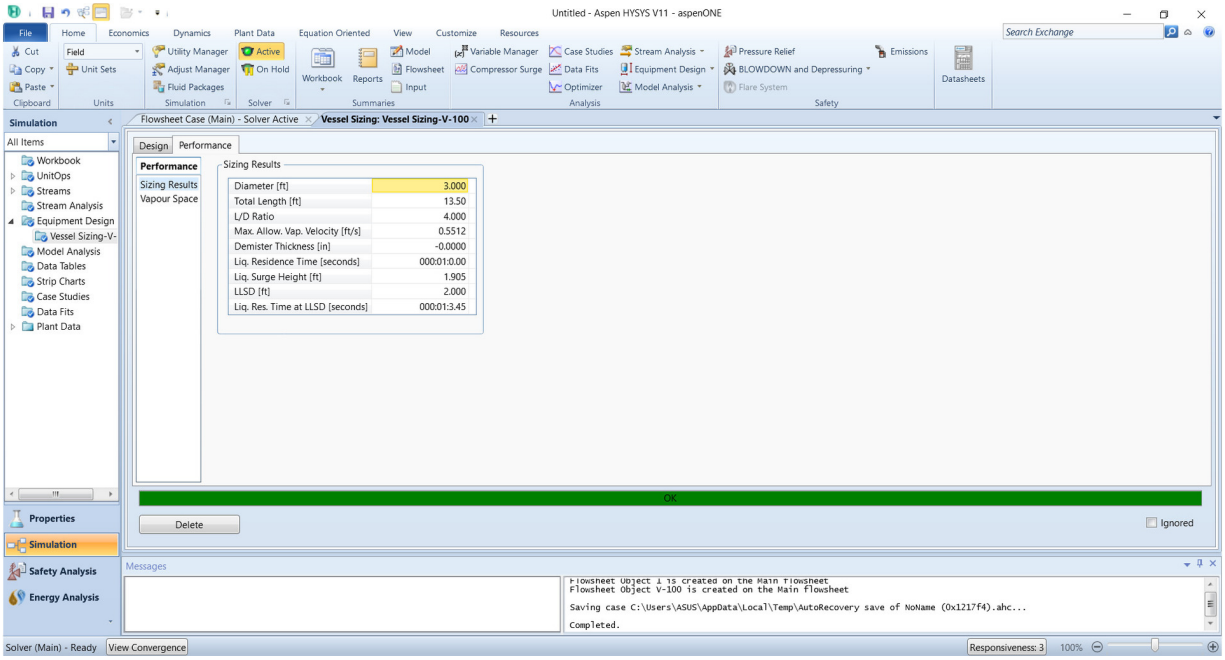
Now let’s choose a smaller diameter for the separator. Diameter is 3 ft.

Figure 9.13: Separator sizing



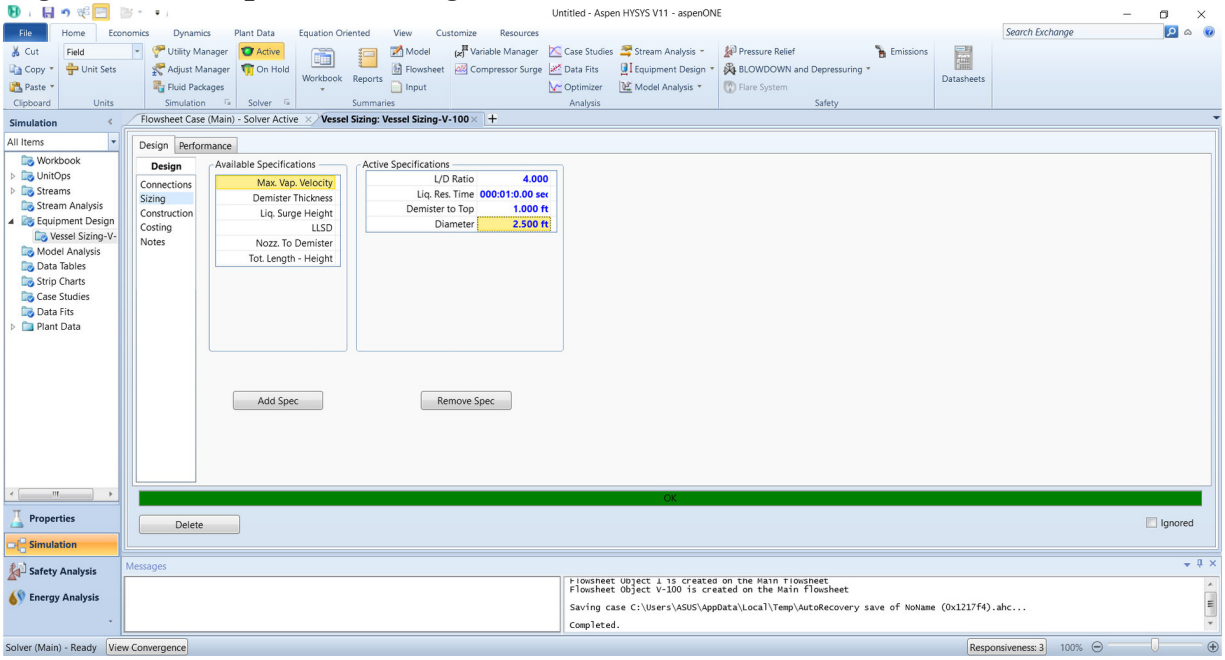
The new height is 13.5 ft and the liquid level is 1.9 ft (14%).

Figure 9.14: Separator sizing



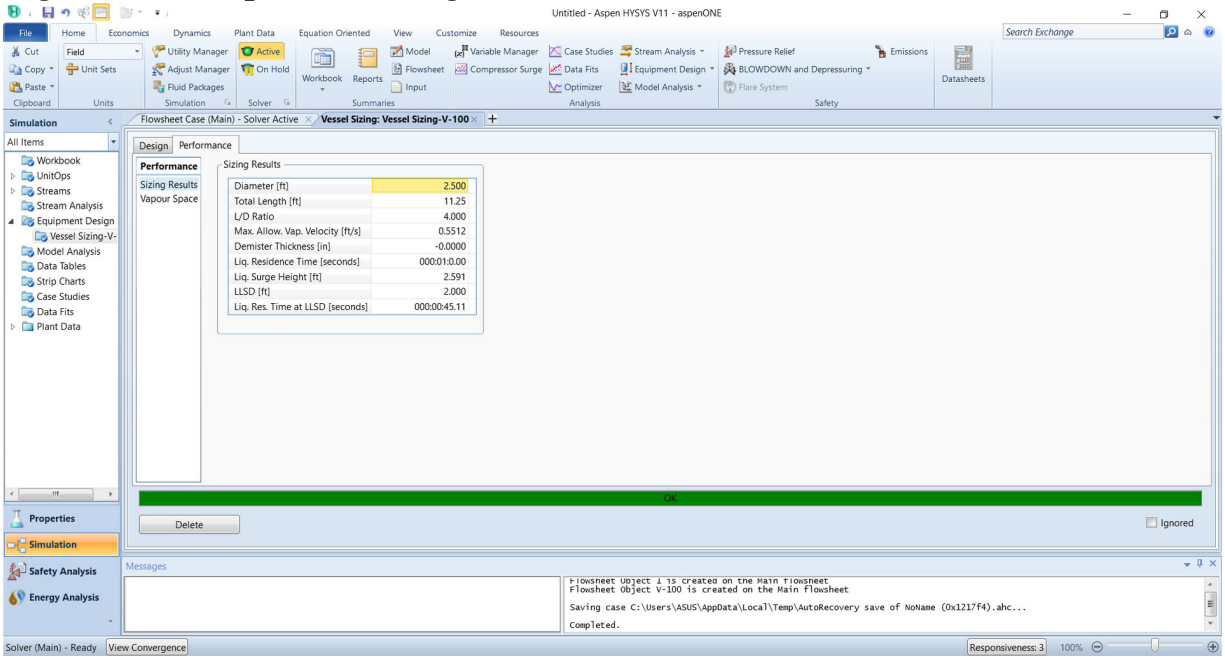
We will further try to optimize the separator with 2.5 ft diameter.

Figure 9.15: Separator sizing



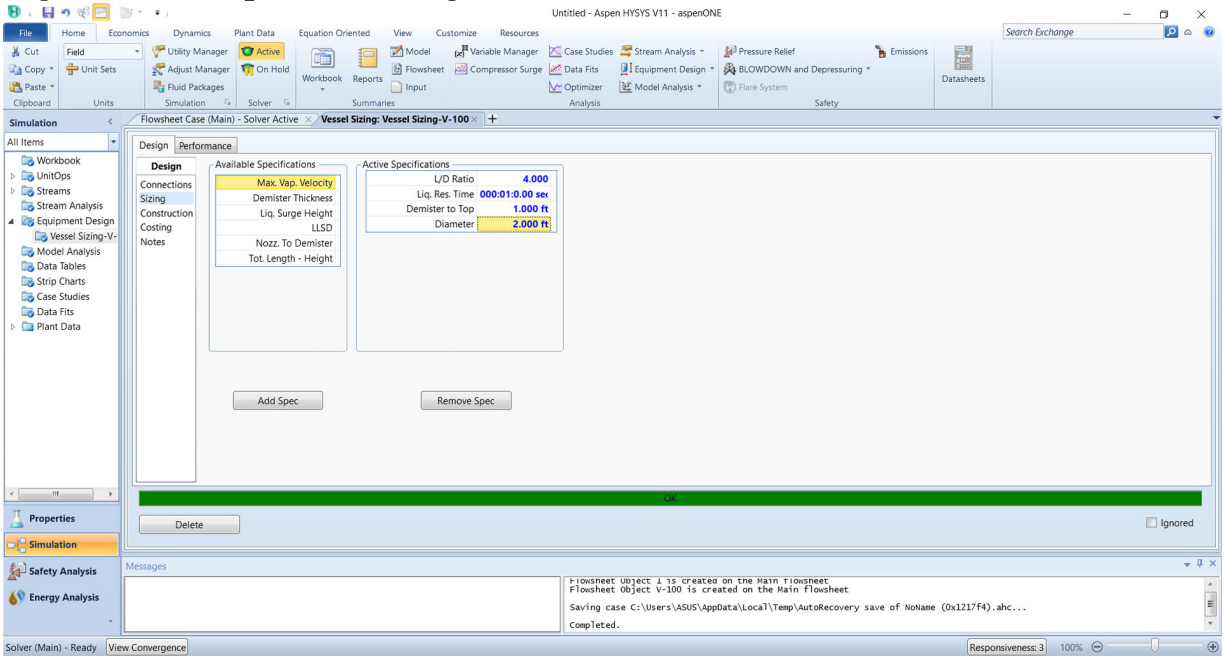
Separator height is 11.25 ft and the liquid level is 2.6 ft (23%).

Figure 9.16: Separator sizing



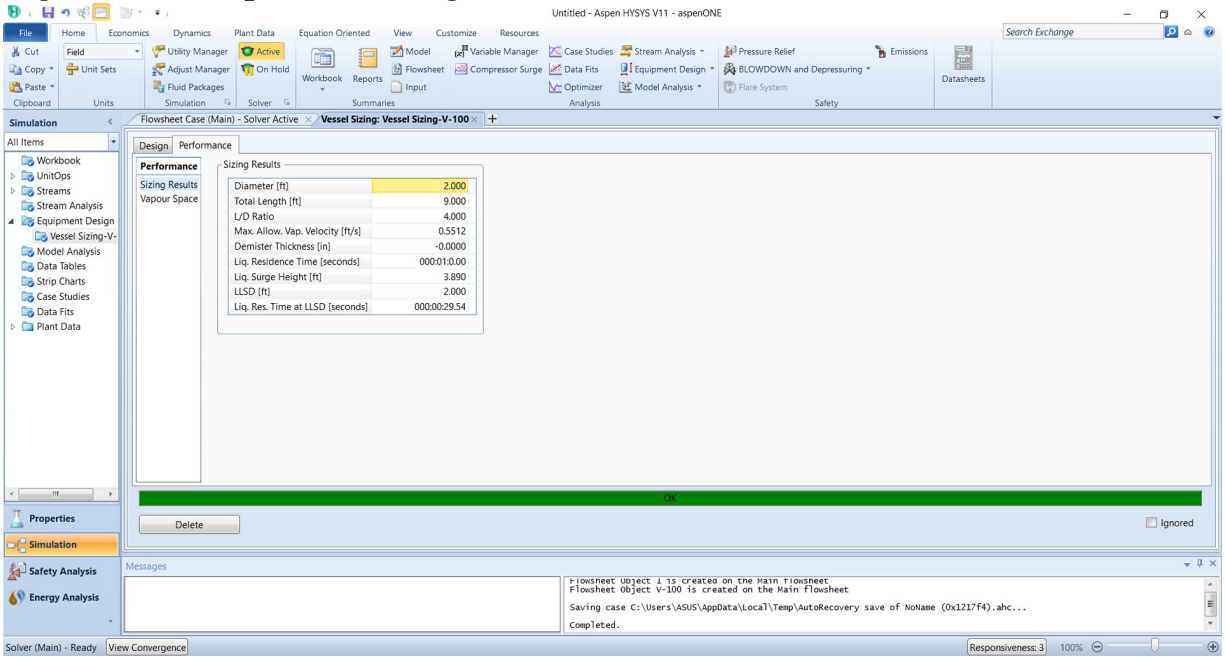
We will further try to optimize the separator with 2.5 ft diameter.

Figure 9.17: Separator sizing



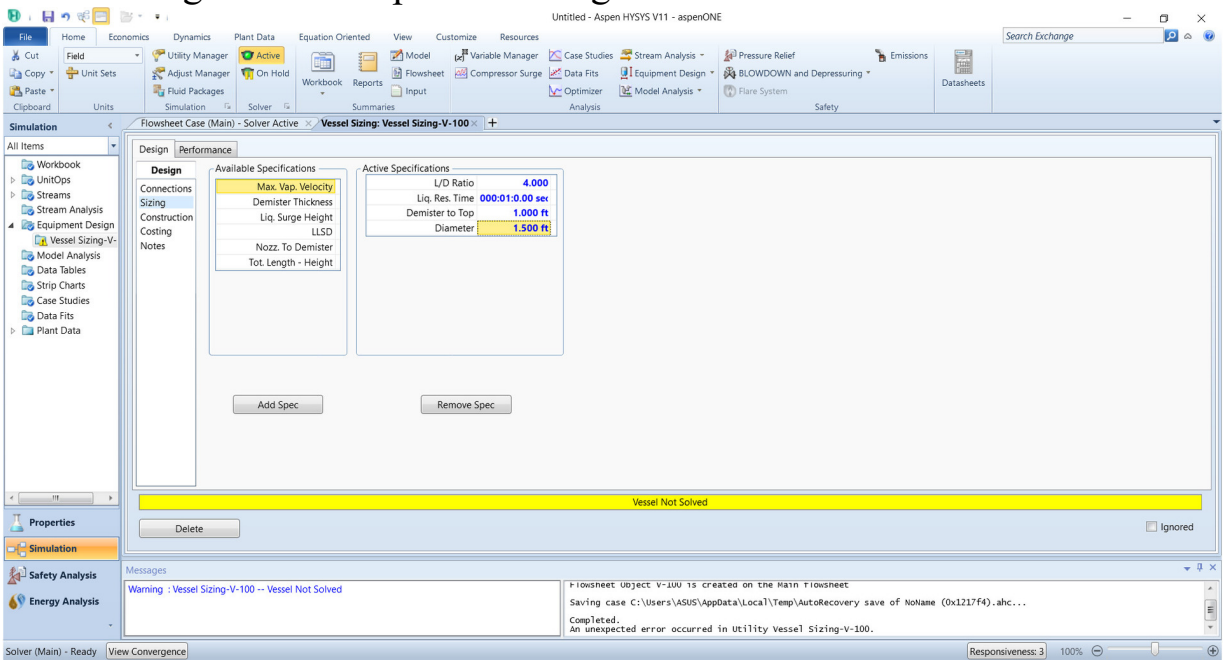
Separator height is 9 ft and the liquid level is 3.9 ft (43%).

Figure 9.18: Separator sizing



But when we further try to reduce the separator diameter, then HYSYS fails to find a solution.

Figure 9.19: Separator sizing



Hence, the solution to the given problem is to select either 2 ft diameter or 2.5 ft diameter separator.

Exercises

a. Ms. Srijita is supposed to discuss separator design. Can you help her?
b. Ms. Shubhangi is trying to size the separator with FEED flow rate 5000 lbmol/h (Methane 81.32% and n-Butane 18.68% by mole), 85 ° F and 900 psia. What would be the appropriate separator size if the flow rate is doubled?

CHAPTER 10: HEAT EXCHANGER SIZING USING ASPEN EDR

Aspen Exchanger Design and Rating (EDR) enables you to find the optimal design for your heat exchanger needs based on cost. The seamless integration between the thermal and mechanical design tools of Aspen EDR and process simulation tools from AspenTech enables analysis of several different alternatives before presenting you with the most optimal design. Use this tutorial to get started using Aspen Shell & Tube Exchanger to create, evaluate, and save designs.

Simulation

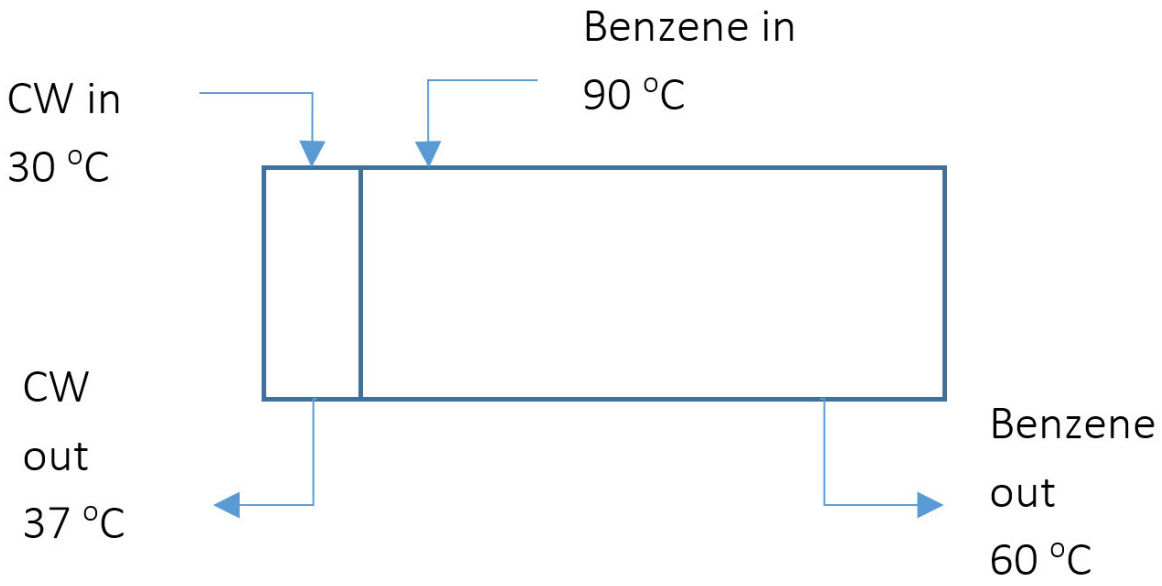
Let's solve an example.

Design a shell and tube exchanger for the following duty 20000 kg/h of benzene. It enters the heat exchanger at 90 ° C and to be cooled to 60 ° C. The cooling medium is cooling water entering the exchanger at 30 ° C and exits at 37 ° C. Benzene inlet pressure is 5 bar and that of water is 6.5 bar.

A pressure drop of 0.8 bar is permissible on both the sides. Allowance should be made for fouling by including a fouling factor of 0.0003 m^2 .K/W on the Benzene side and 0.0001 m^2 .K/W on waterside.

We limit the exchanger length as 5 m.

Figure 10.1: Heat exchanger problem



SOLUTION

Open Aspen EDR new file. Check the “Shell & Tube” box and “Create” the new file.

Figure 10.2: New case

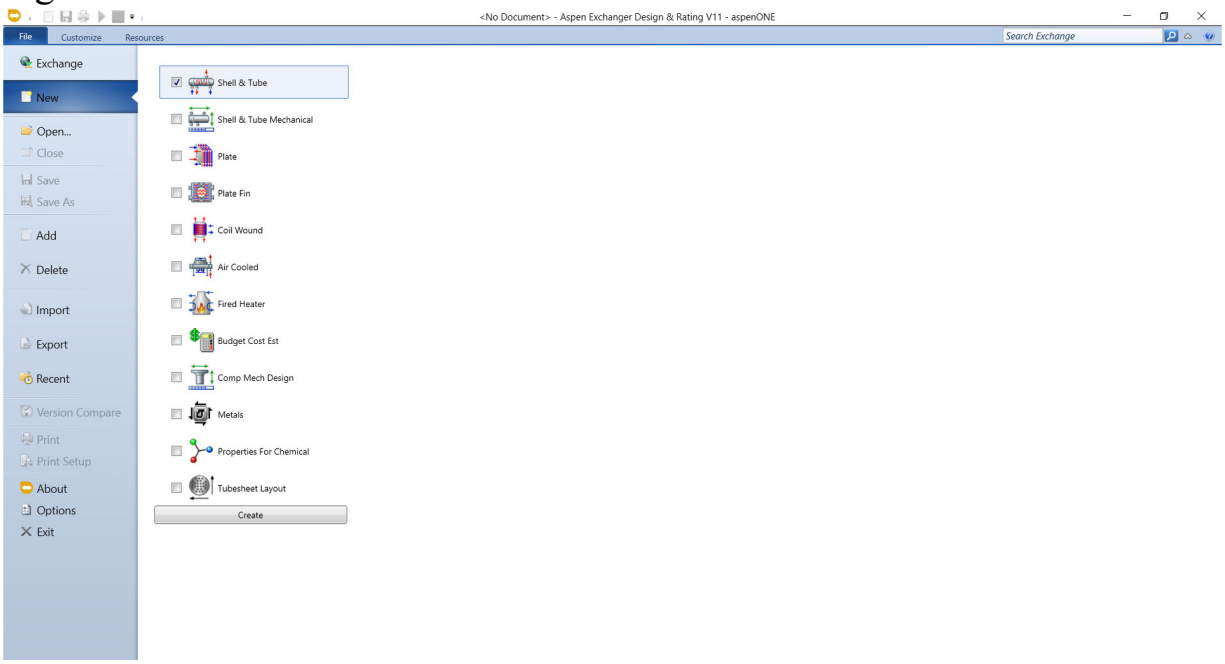
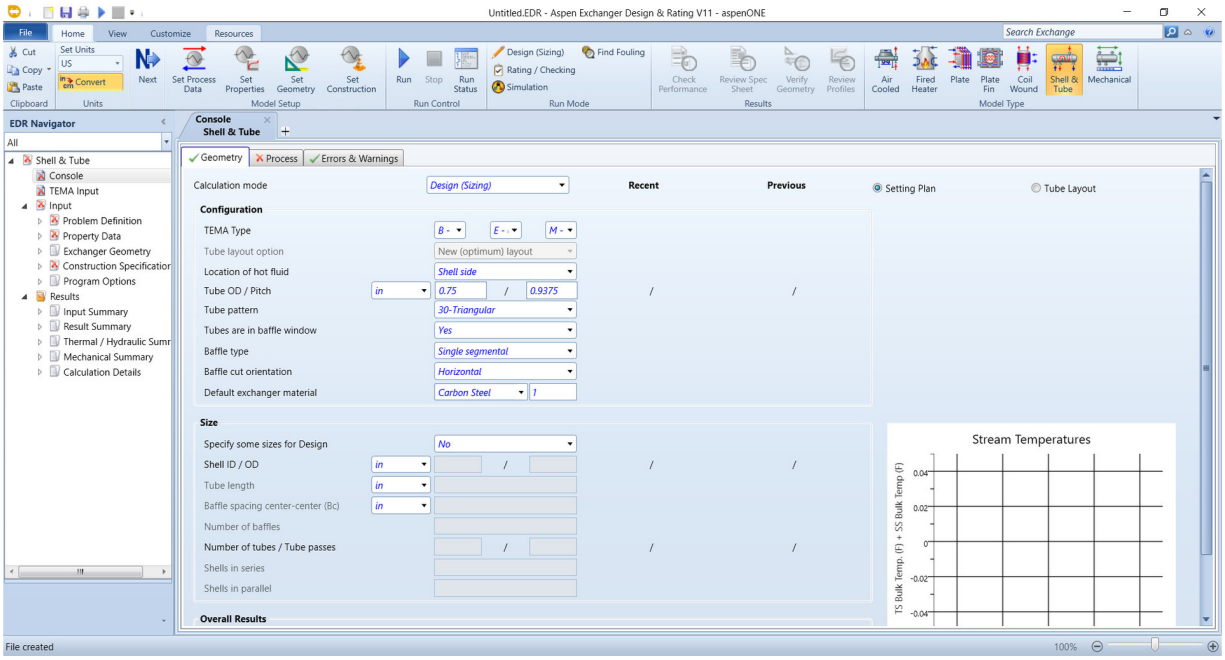
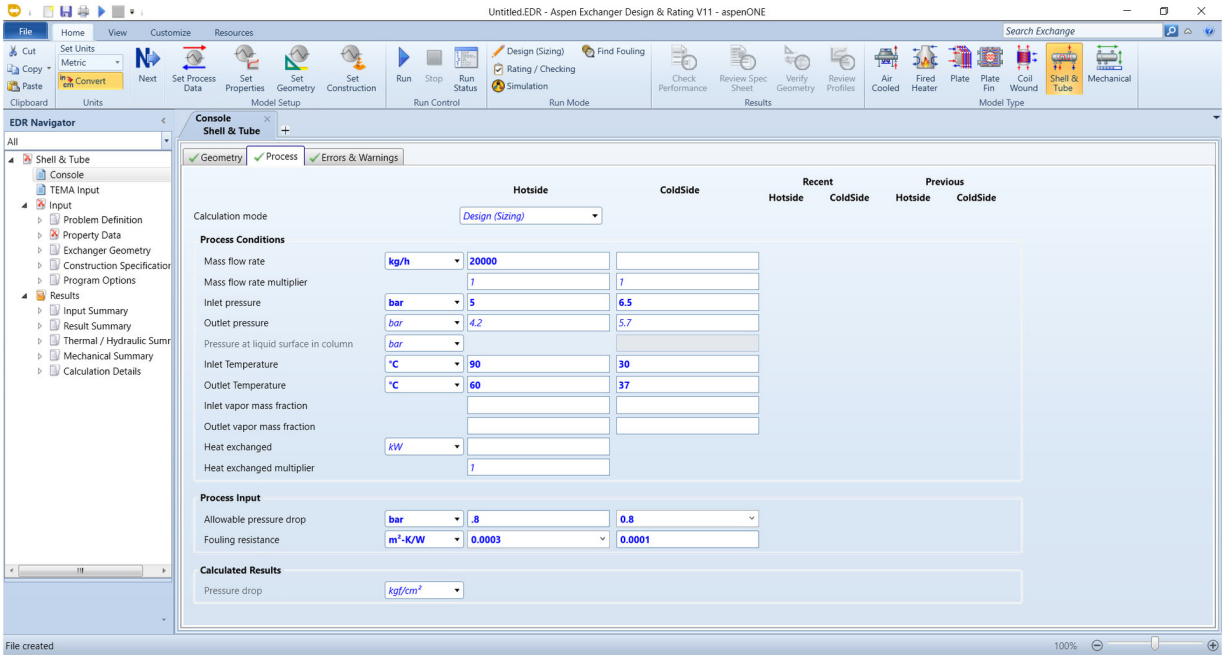


Figure 10.3: New case



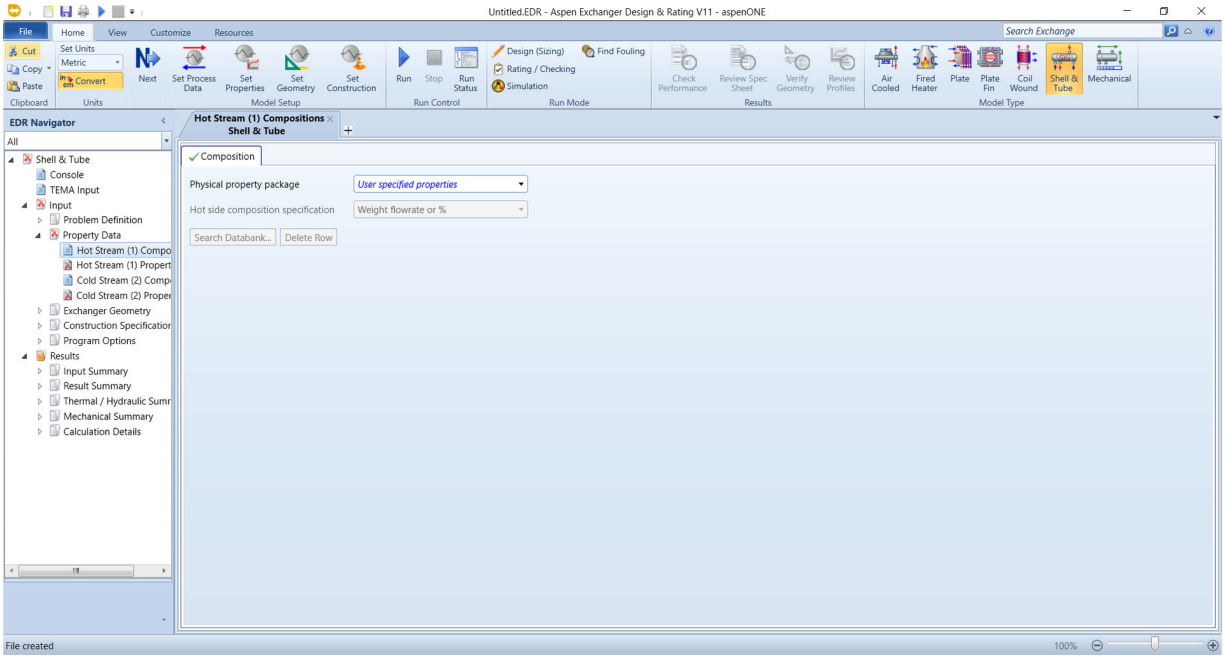
Provide available data (flow rate, pressure, pressure drop, temperature, and fouling resistance) in the “Process” tab.

Figure 10.4: EDR



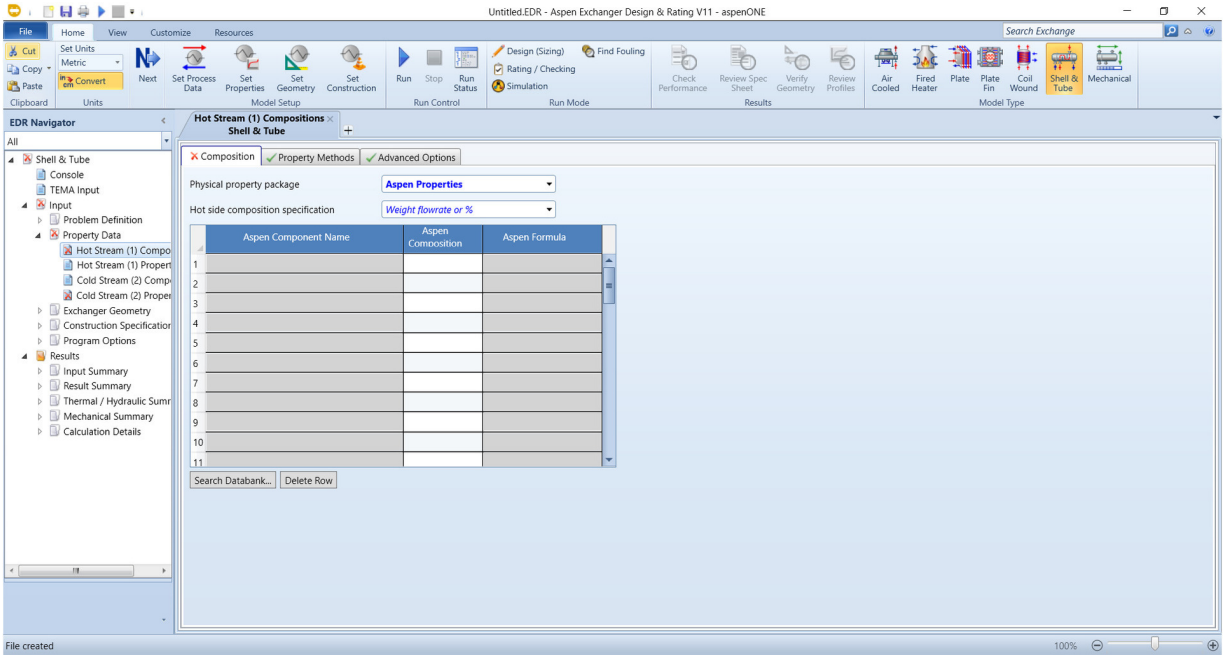
Select “Property Data\Hot Stream (1) Composition”.

Figure 10.5: Composition



Select “Physical property package” as “Aspen Properties”.

Figure 10.6: Property package



Click on “Search Databank” and “Find Now” Benzene and Water. You need to “add selected compounds”.

Figure 10.7: Components

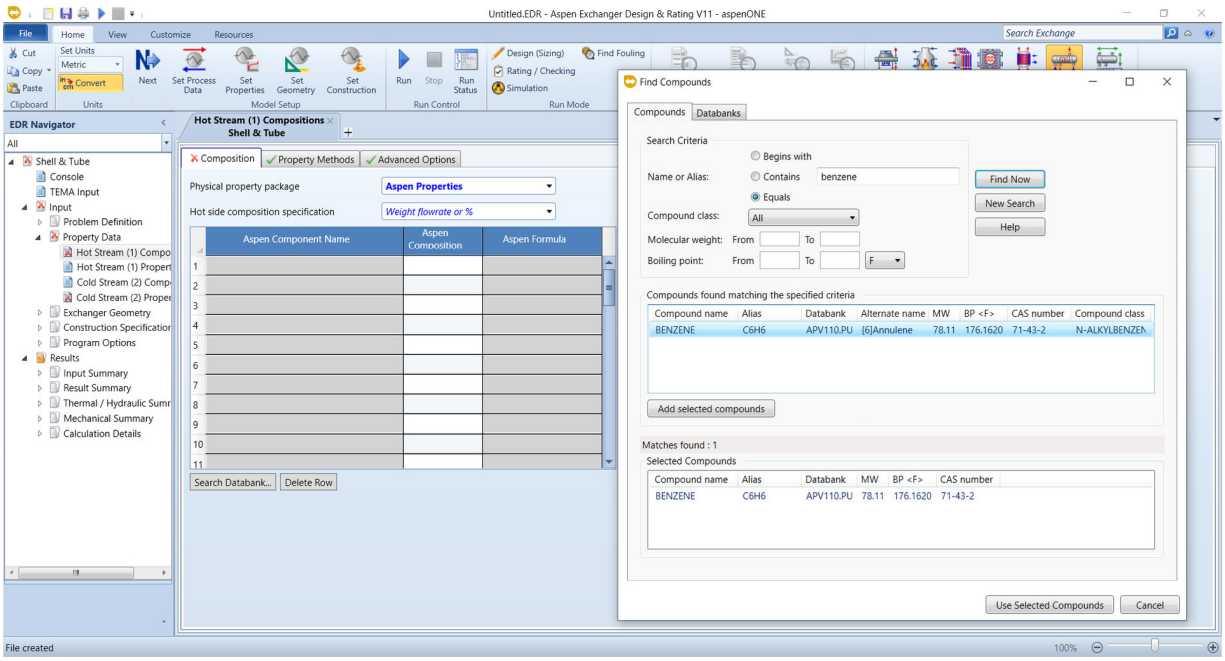
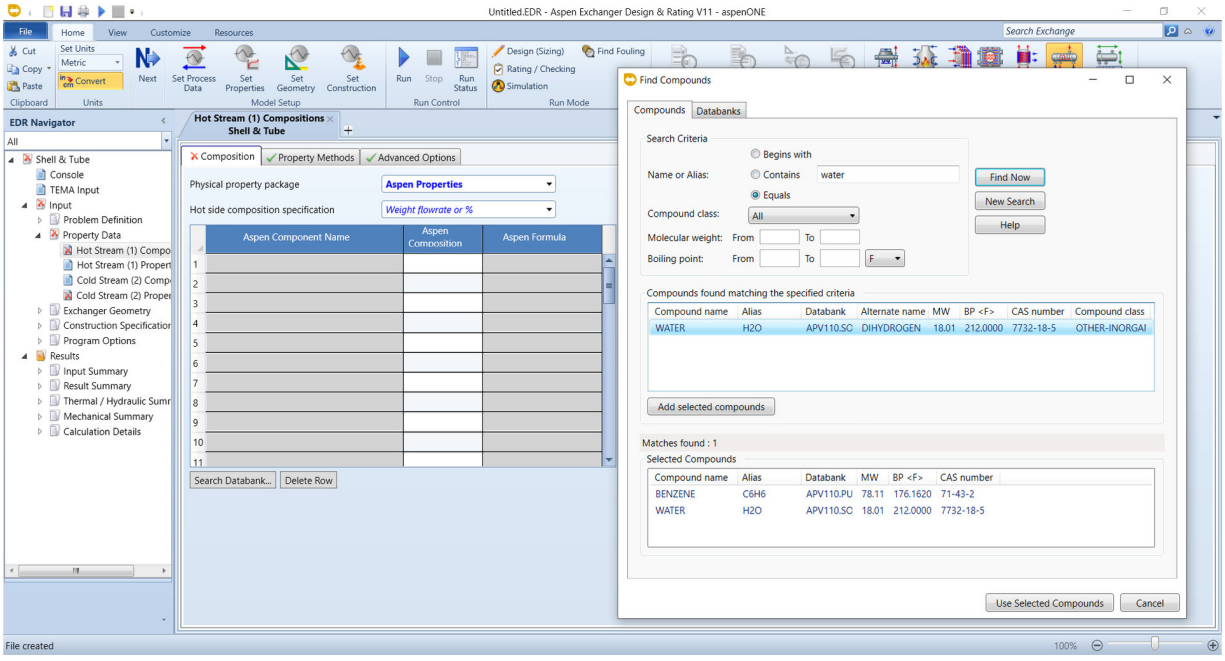
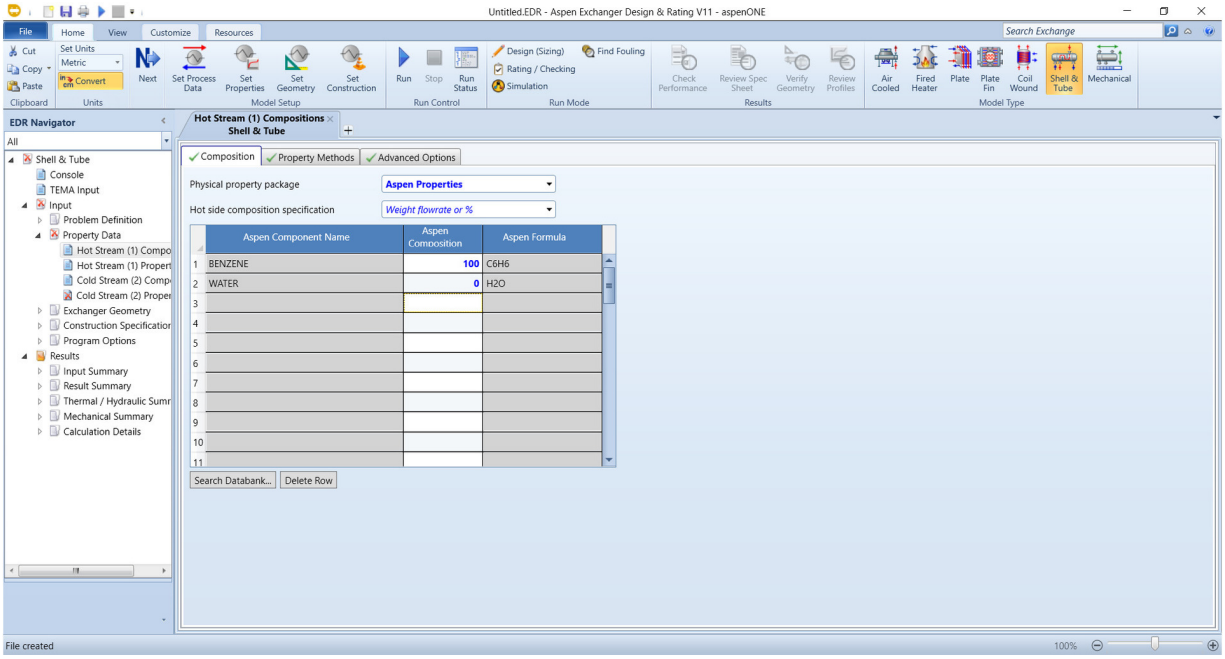


Figure 10.8: Components



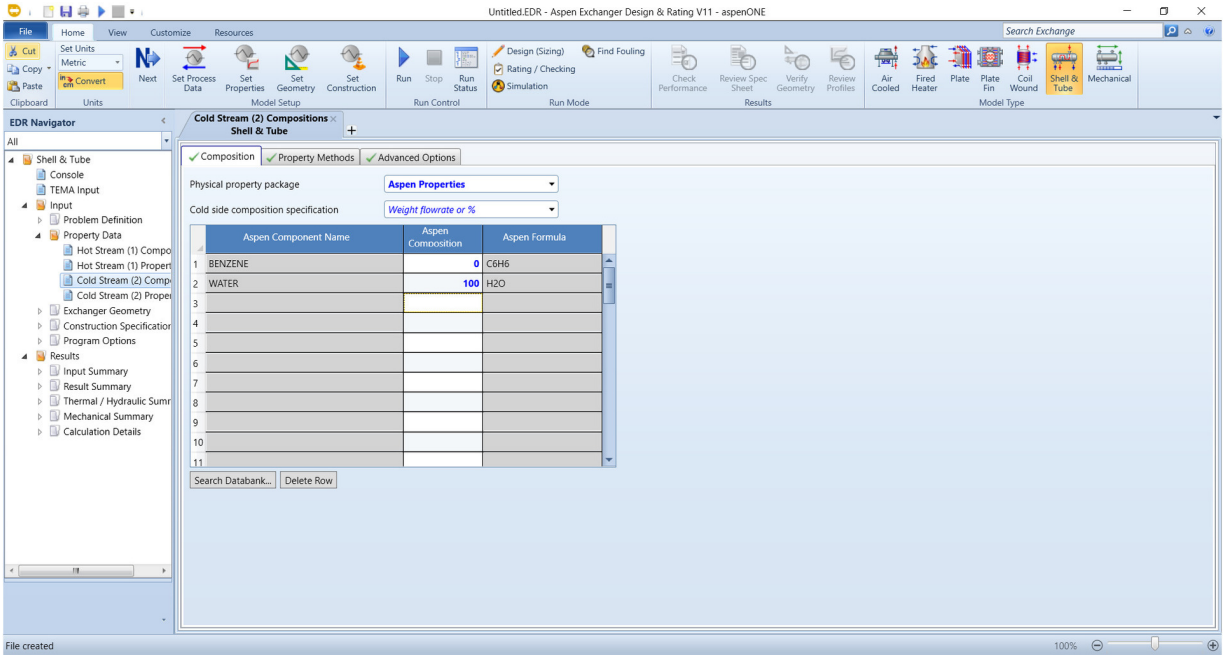
Provide Weight percentage for a hot stream (Benzene) as 100% Benzene and 0% Water.

Figure 10.9: Components



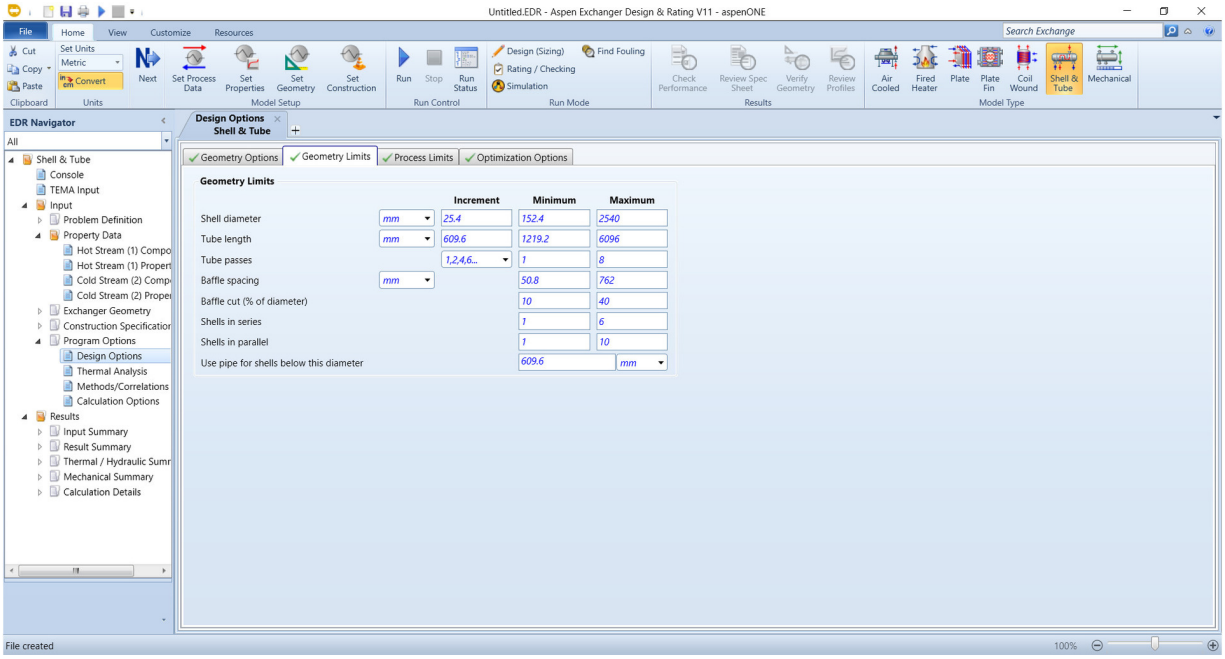
The same procedure shall be repeated for “Cold Stream”.

Figure 10.10: Components



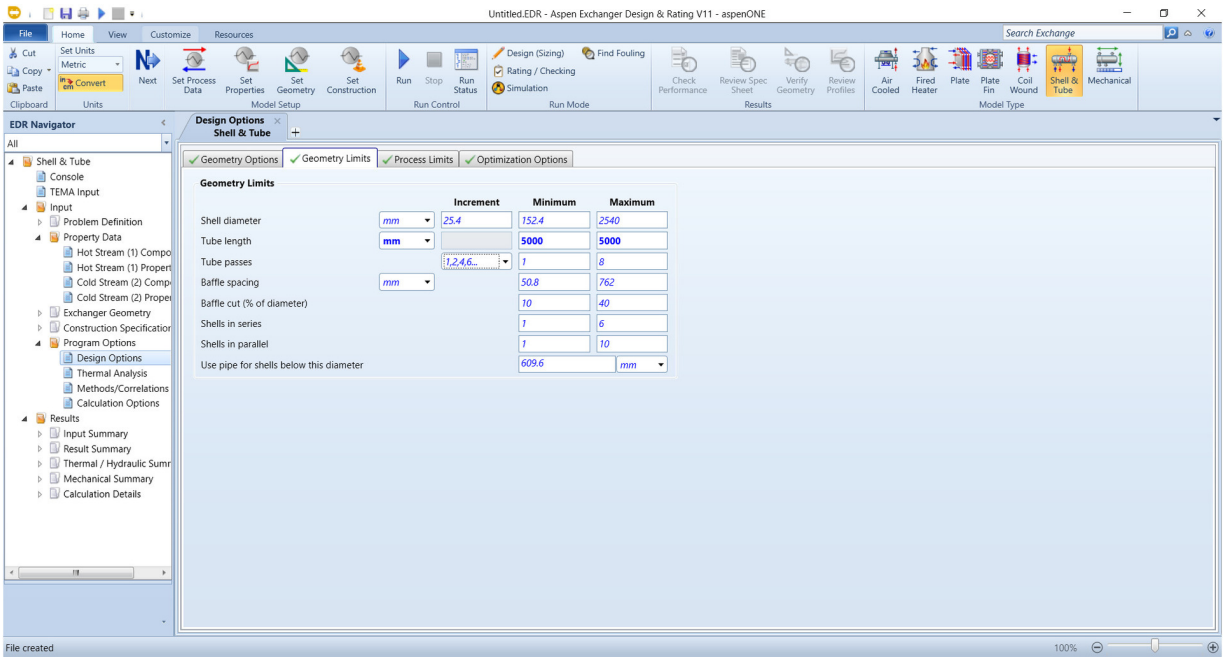
Since the problem statement has a constraint that the tube length should be 5 m, the same shall be provided in "Program Options\Design Options\Geometry Limits".

Figure 10.11: Design limits



Tube length shall be provided as 5000 mm minimum and 5000 mm maximum.

Figure 10.12: Design limits



Click “Run” or press F5. Aspen EDR will Design the heat exchanger as per input parameters. The results are summarized in “Results\Result Summary\TEMA Sheet”.

Following can be noted from TEMA sheet;

- Heat transfer area = 12.4 m^2
- CW required = 33218 kg/h
- Heat exchanged = 250555 kcal/h
- LMTD = 39.7 ^{o} C
- U_D = 509.8 kcal/(h.m^2 .^{o} C)
- No. of tube = 42

- No. of tube passes = 2
- Shell ID = 205 mm

Figure 10.13: TEMA sheet

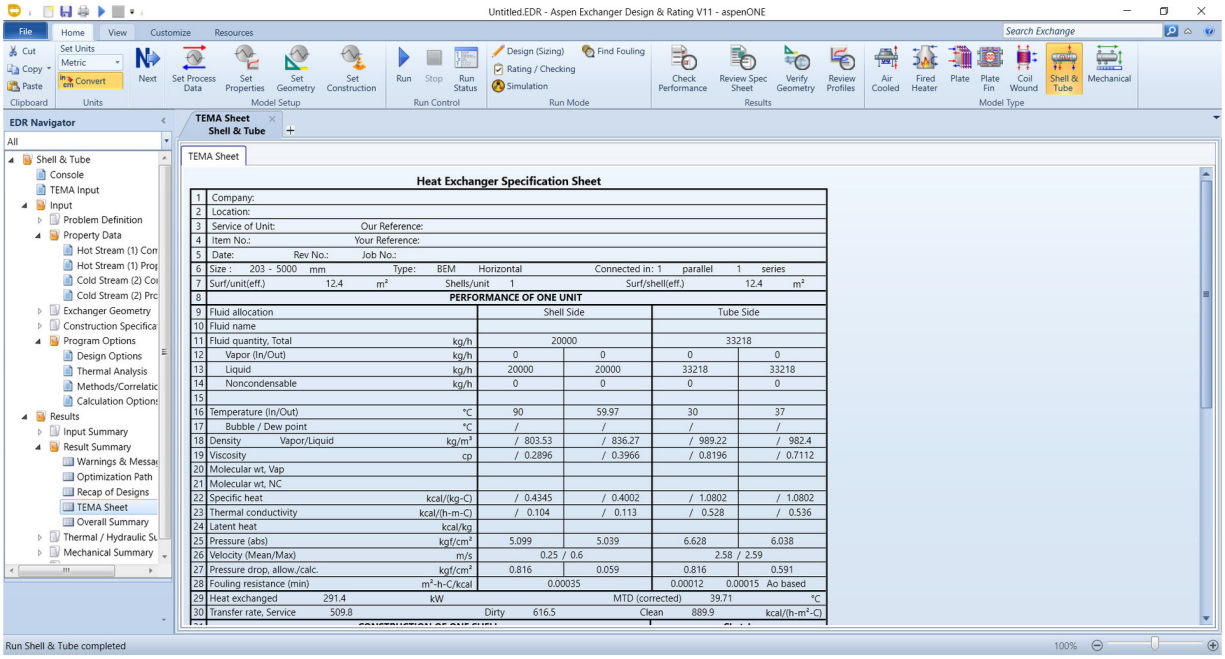
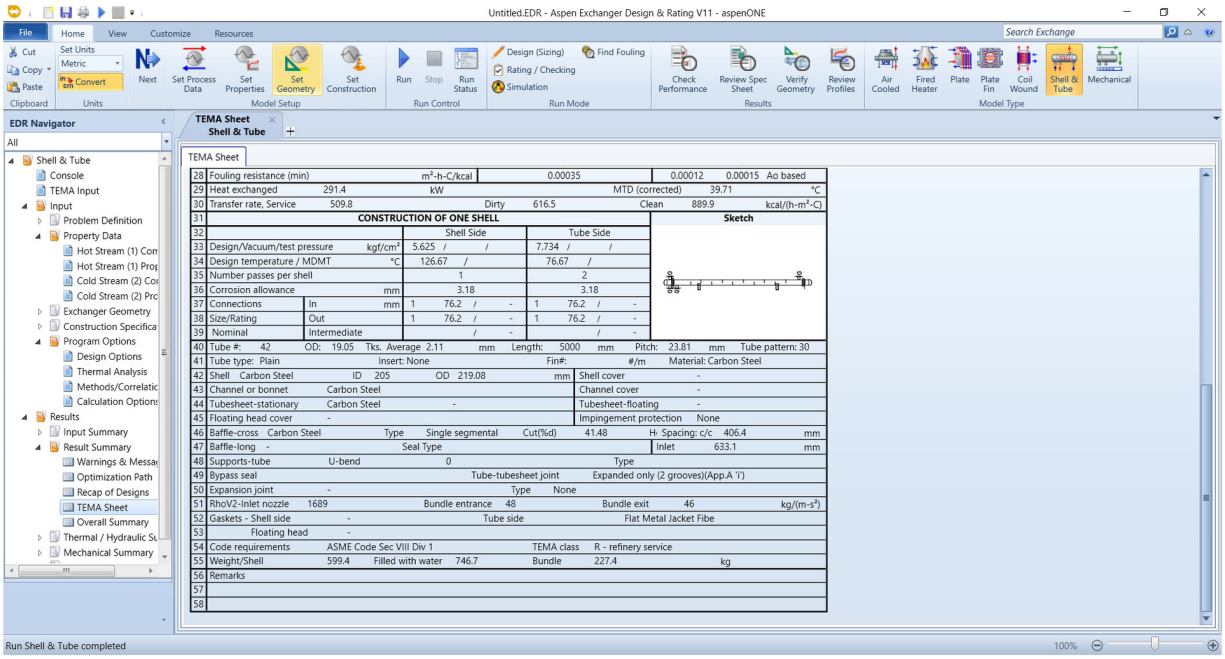
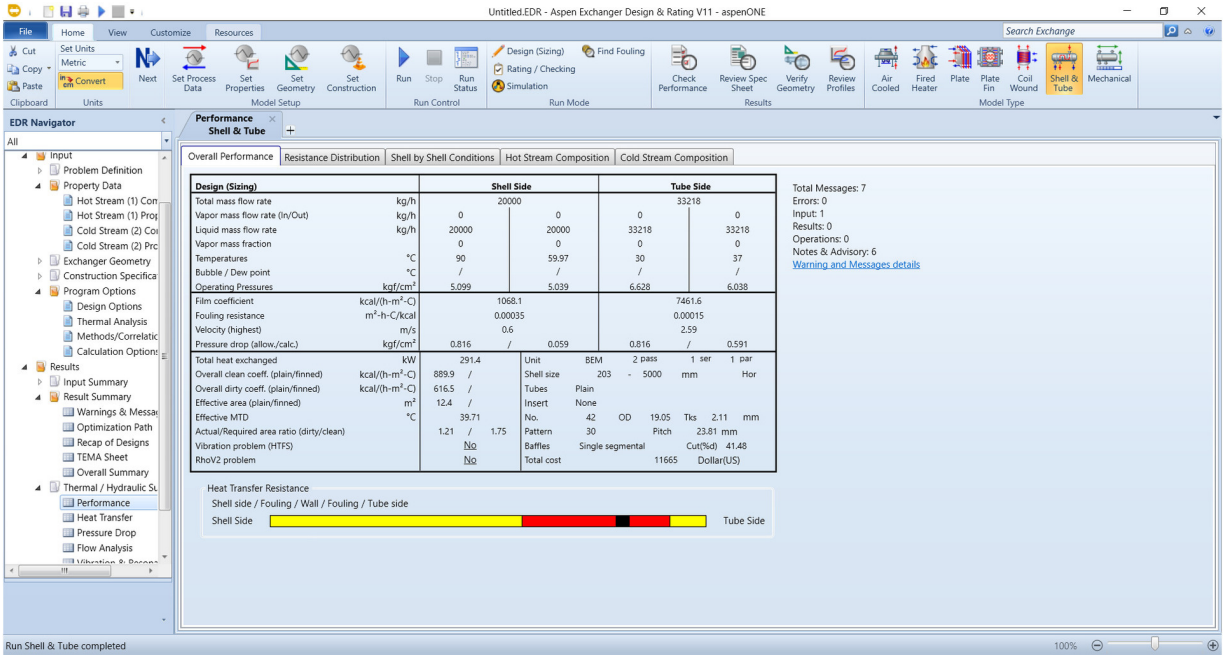


Figure 10.14: TEMA sheet



From the “Thermal/Hydraulic Summary\Performance” tab, it is evident that shell-side fluid (Benzene) has more resistance to heat transfer compared to the Tube side (cooling water).

Figure 10.15: Performance



“Setting plan and Tubesheet layout” is given below.

Figure 10.16: Settling plan

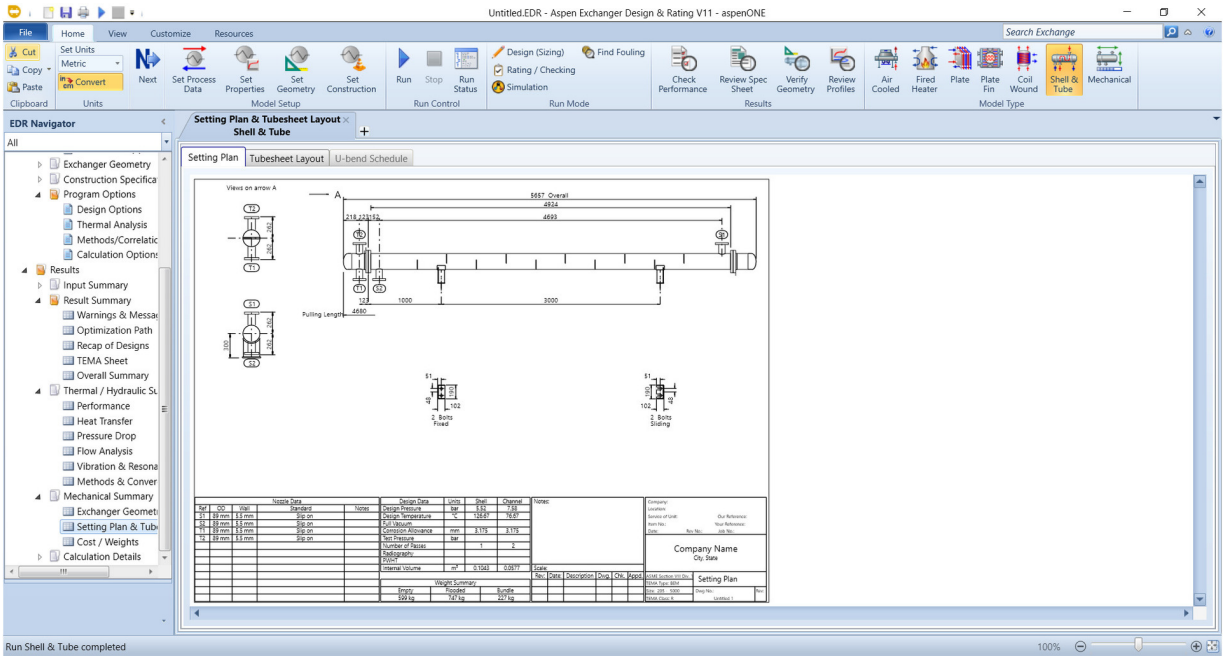
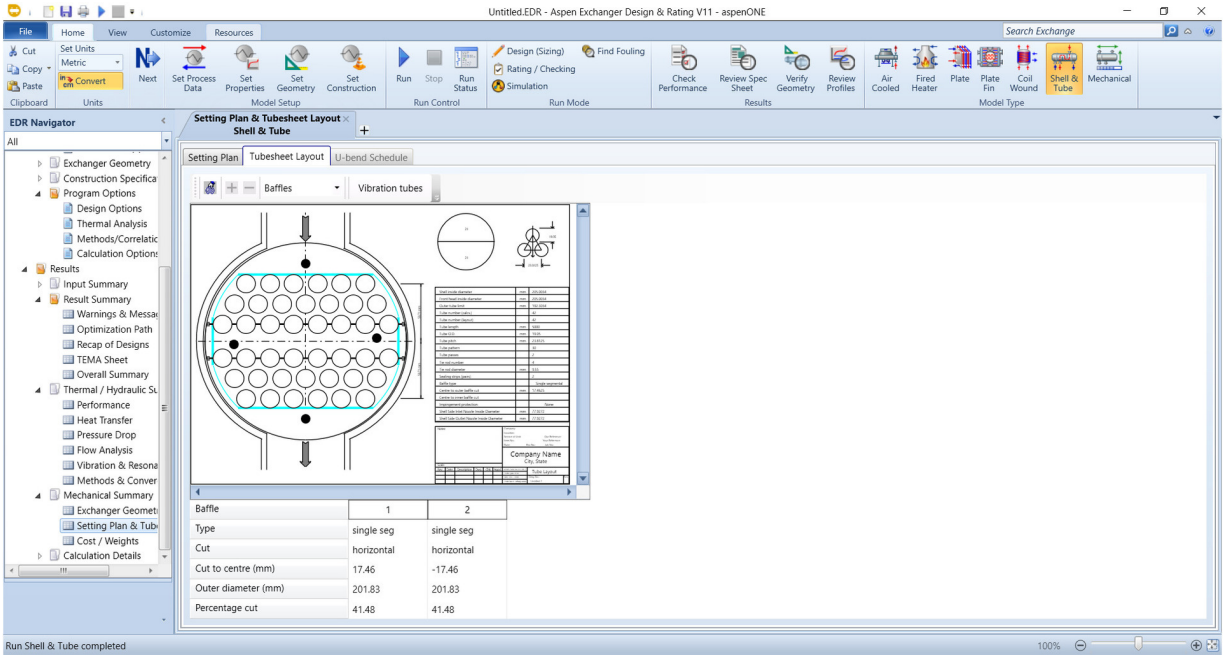
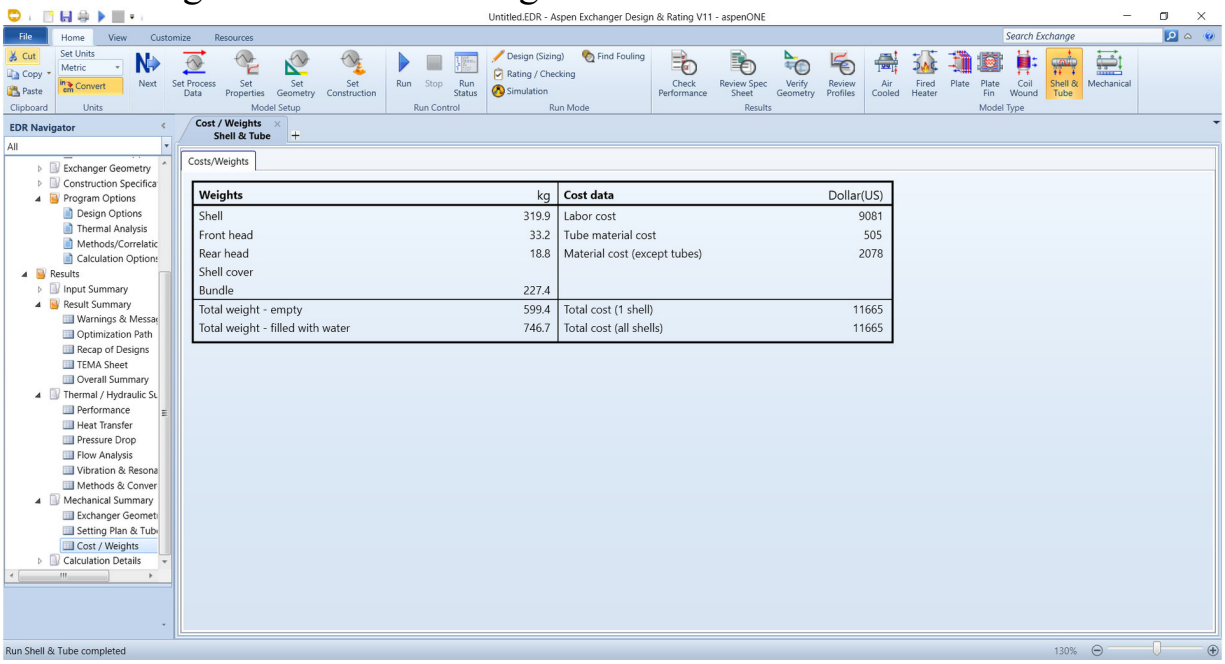


Figure 10.17: Settling plan



"Cost/Weight" data is also calculated by Aspen EDR. Weight result is, however, more reliable than the cost.

Figure 10.18: Cost/Weight



Exercises

a. Mr. Sai works in the technical services department in an operating company. For a new project, he is to design a heat exchanger with the following data. Design a shell and tube exchanger for the following duty 40000 kg/h of benzene. It enters the heat exchanger at 100 ° C and to be cooled to 65 ° C. The cooling medium is cooling water entering the exchanger at 30 ° C and exits at 35 ° C. Benzene inlet pressure is 5.5 bar and that of water is 5 bar. A pressure drop of 1 bar is permissible on both sides. Allowance should be made for fouling by including a fouling factor of 0.0002 m^2 .K/W on the Benzene side and 0.0001

m^2 .K/W on waterside. What heat exchanger size he would have calculated?

b. Mr. Sachin is Mr. Sai's boss and has chosen to disregard his design and instead used a 50% oversize heat exchanger. How do you think the exchanger would perform in such a case?
c. Ms. Shweta believes that if tube length is considered as 6 m, then the heat transfer area would be minimum. Can you prove her wrong?
d. Mr. Sai's design was implemented finally. After 6 months, Mr. Nikhil (shift in-charge) realized that the benzene flow rate was to be reduced to 30000 kg/h if he wanted to maintain the temperature out of Benzene. Can you explain this?
e. Prepare a spreadsheet to calculate HTA with no phase change system. Compare the results with the Aspen EDR TEMA sheet.

CHAPTER 11: DISTILLATION COLUMN SIZING USING ASPEN HYSYS

Separation of light products is present in any Hydrocarbons operations. In this chapter, a column will be modeled to separate Light and heavy components from each other using a distillation column with 12 trays.

We need to separate a mixture of five paraffins into light and heavy fraction by using a distillation column with 12 trays, a full reflux condenser, and a Kettle reboiler. The feed stream (2000 lbmol/hr) consists of 20% (mole %) ethane, 20% propane, 20% n-butane, 20% n-pentane, and 20% n-hexane at 255 ^{o} F and 270 psia, which enters the column on the sixth tray, counting from the top. The condenser and reboiler pressures are 255 and 260 psia, respectively.

The preliminary design specifications require a reflux ratio of 2 and a vapor overhead product of 700 lbmol/hr. Subsequently, the design is modified to ensure propane overhead flow of 350 lbmol/hr and n-butane bottom flow of 380 lbmol/hr.

Use SRK Fluid Pkg.

Calculate the following;

- The Condenser, the Reboiler Temperatures & the Reflux Ratio after modification
- Condenser Temp
- Reboiler Temp
- Reflux Ratio

Define the components and Fluid package (SRK).

Figure 11.1: Components

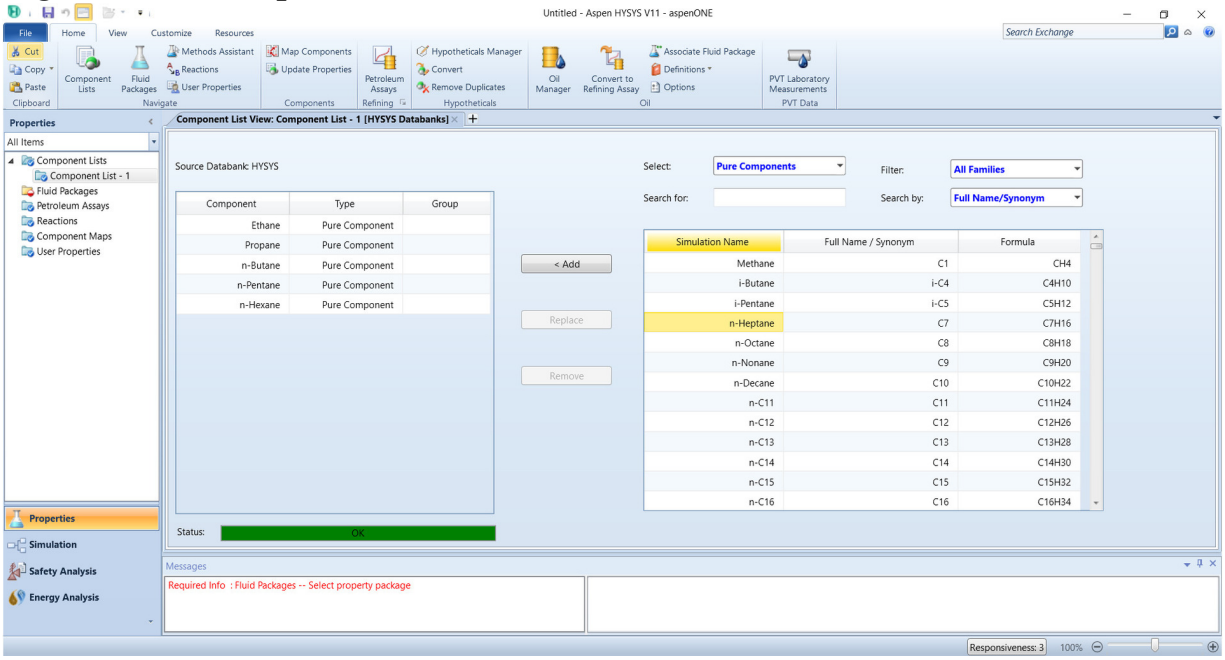
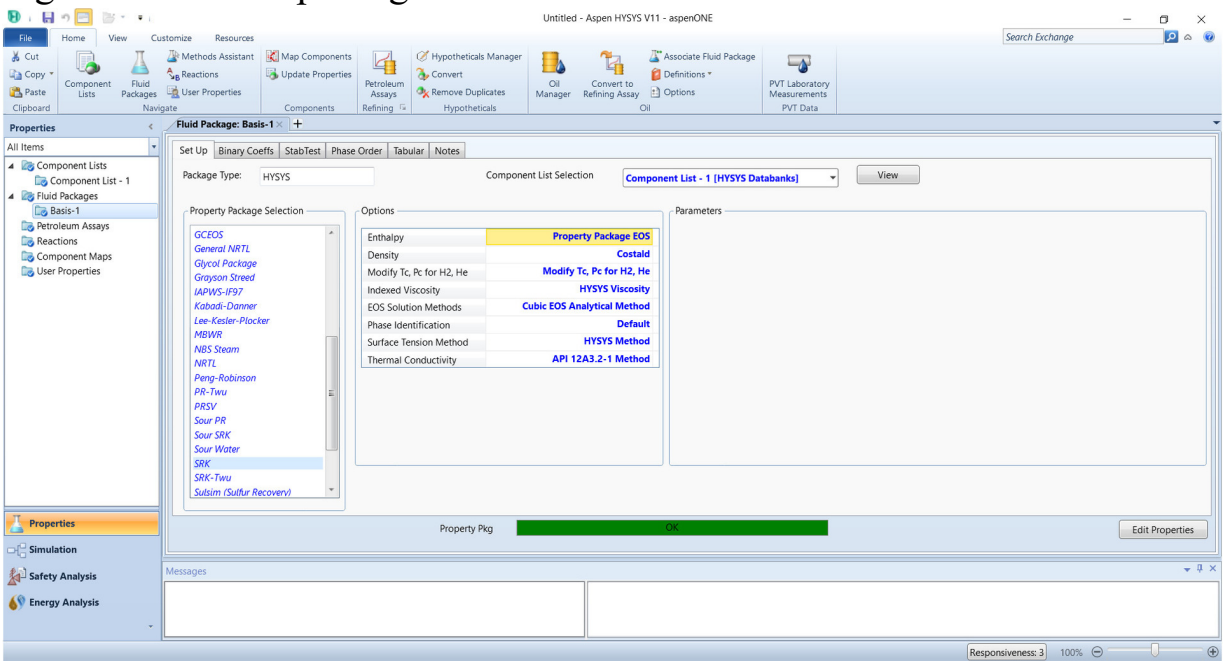


Figure 11.2: Fluid package



Enter the simulation environment and define material stream “1” as below.

Figure 11.3: Material stream

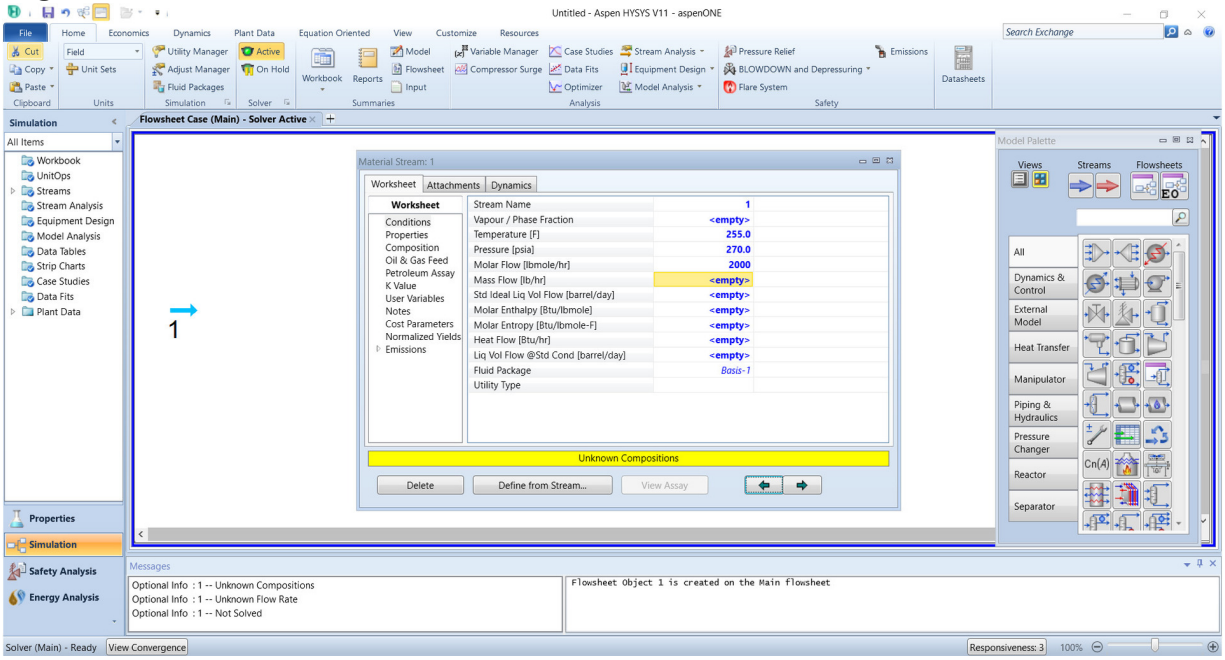
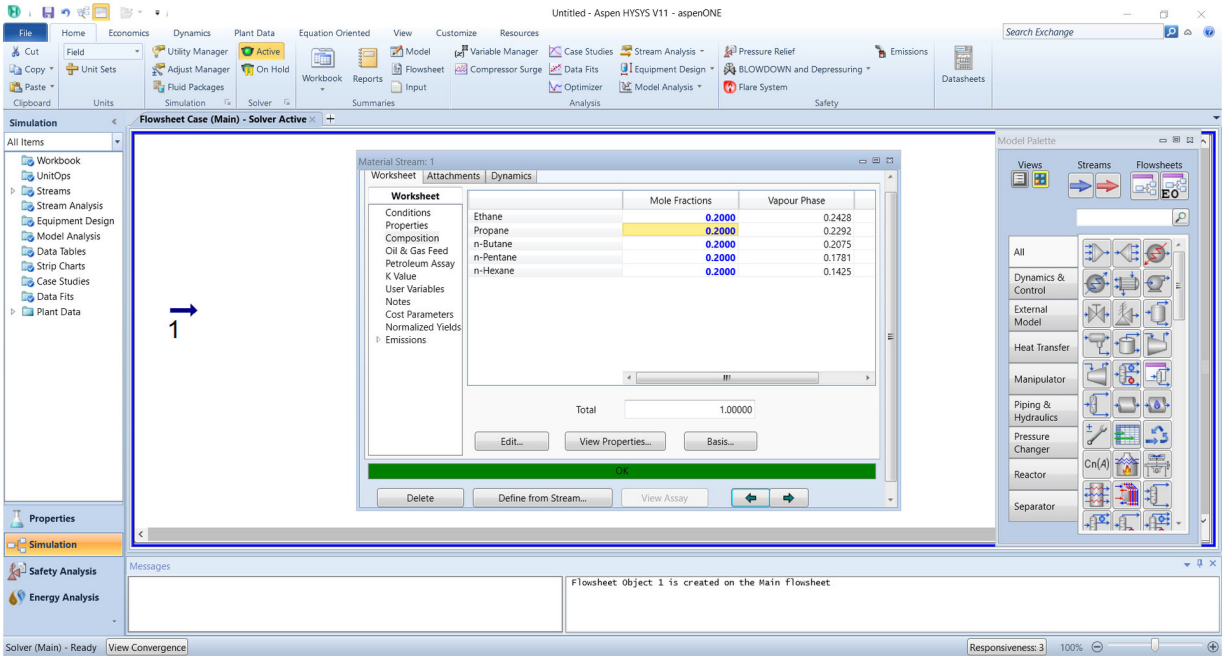
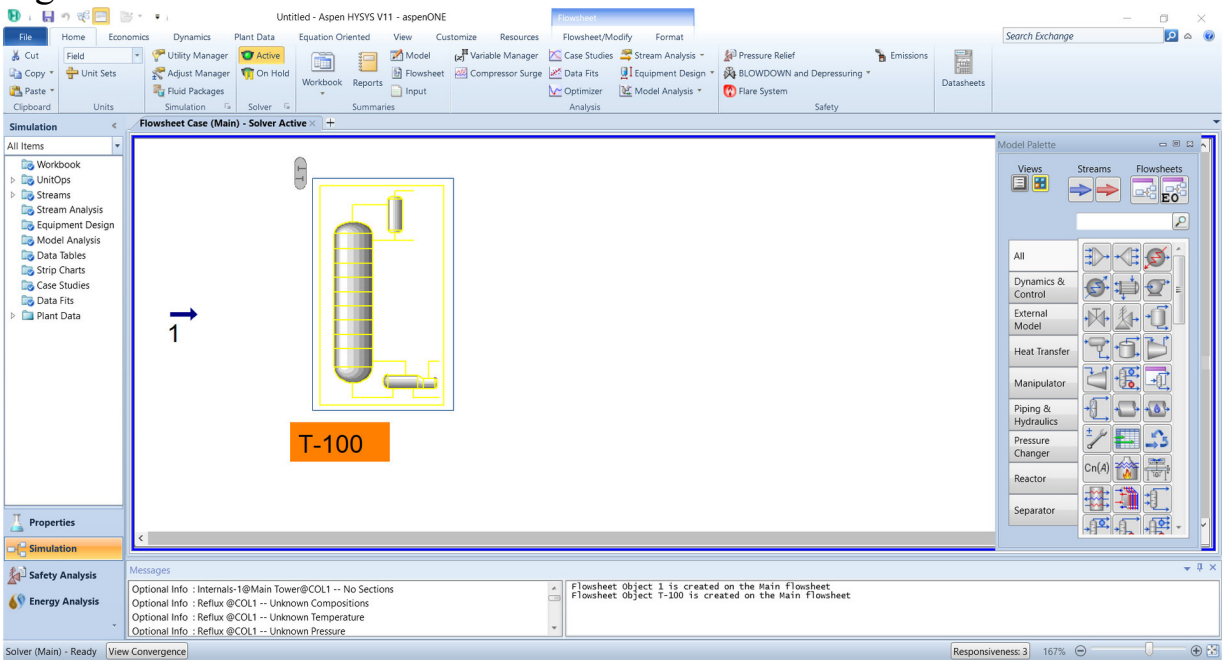


Figure 11.4: Material stream



Add a distillation tower (with condenser & Reboiler).

Figure 11.5: Distillation column



Double-click on T-100.

Figure 11.6: Distillation column

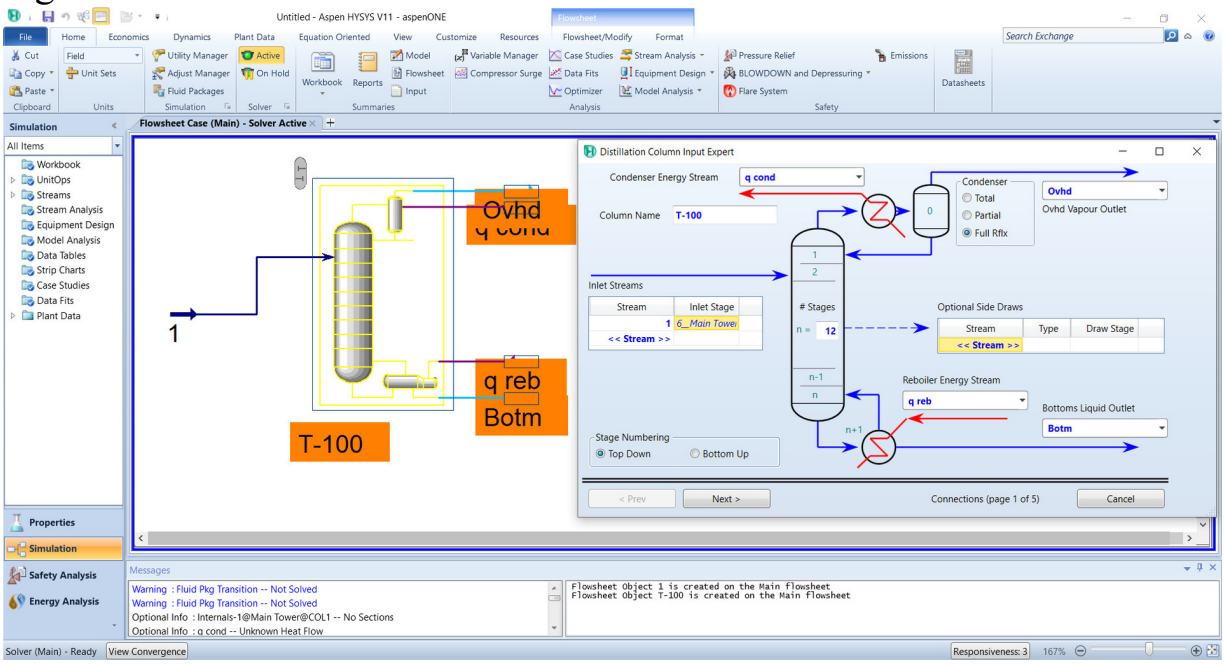


Figure 11.7: Distillation column

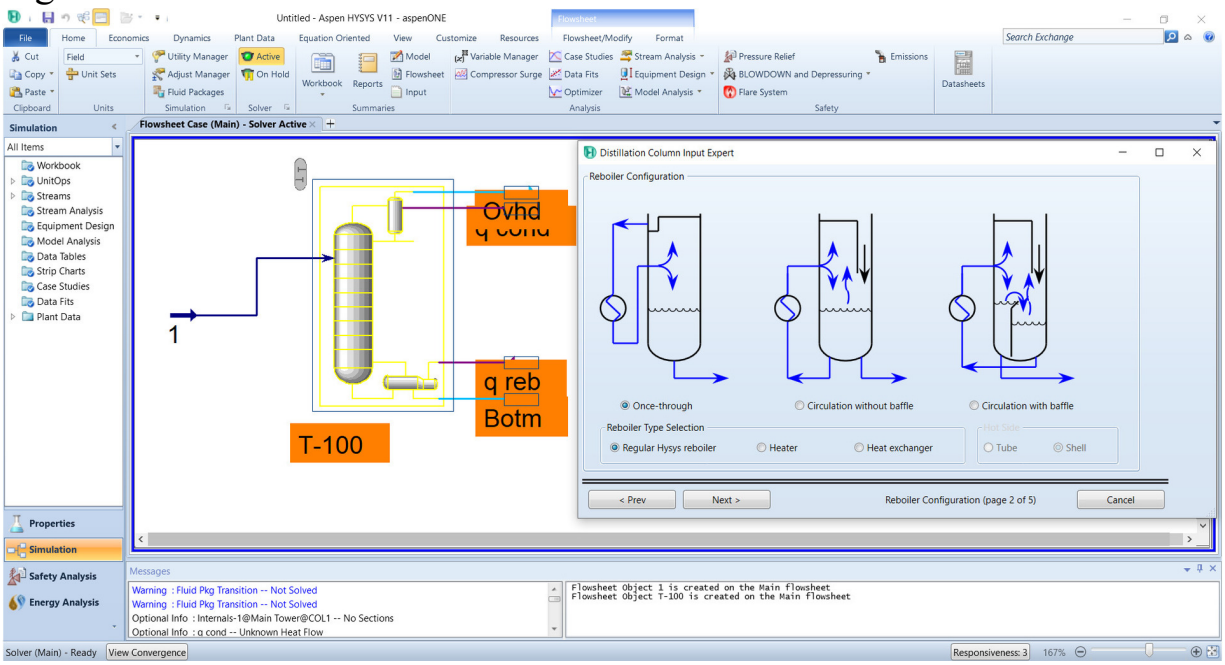


Figure 11.8: Distillation column

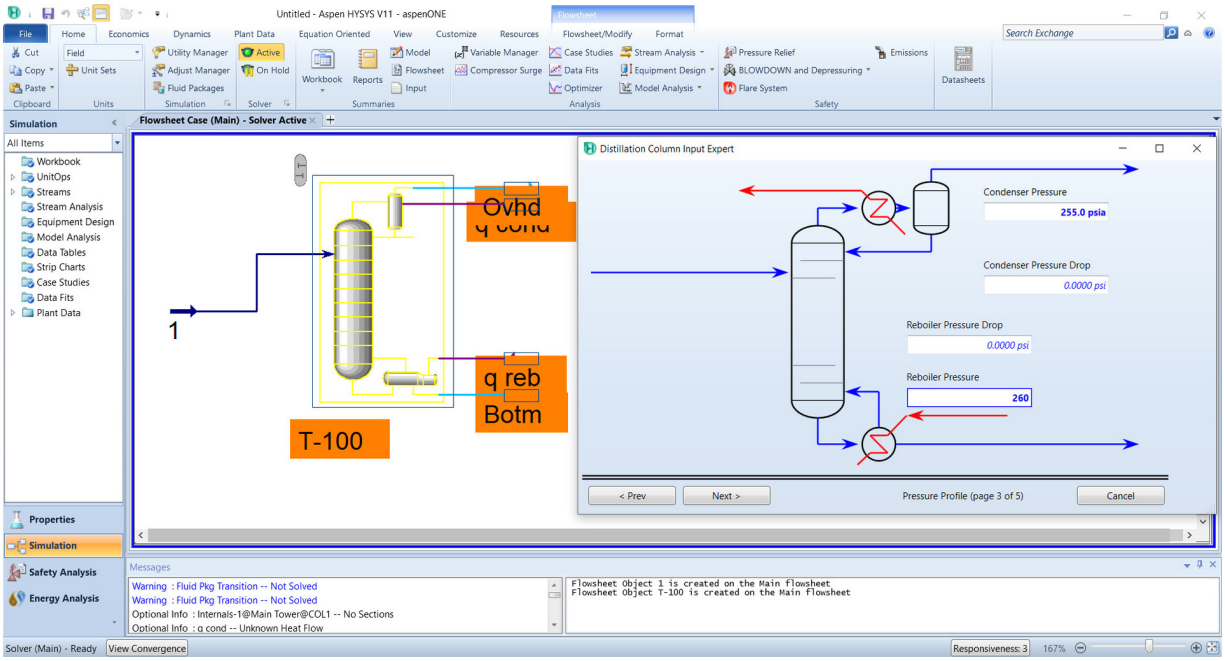


Figure 11.9: Distillation column

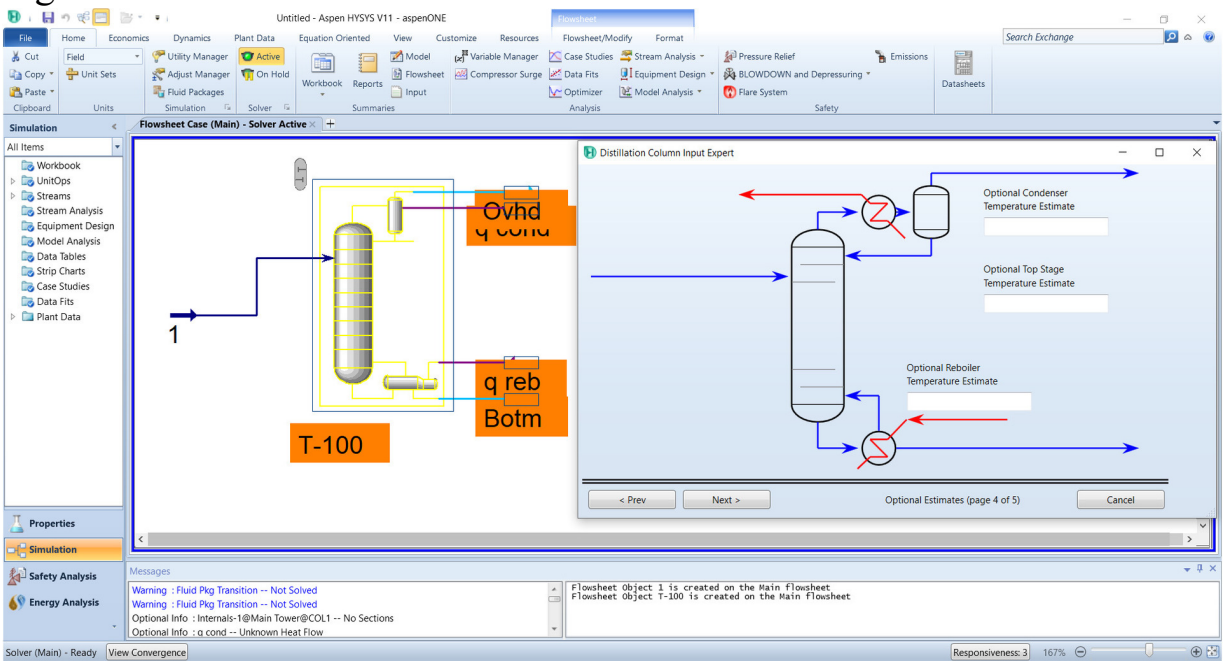


Figure 11.10: Distillation column

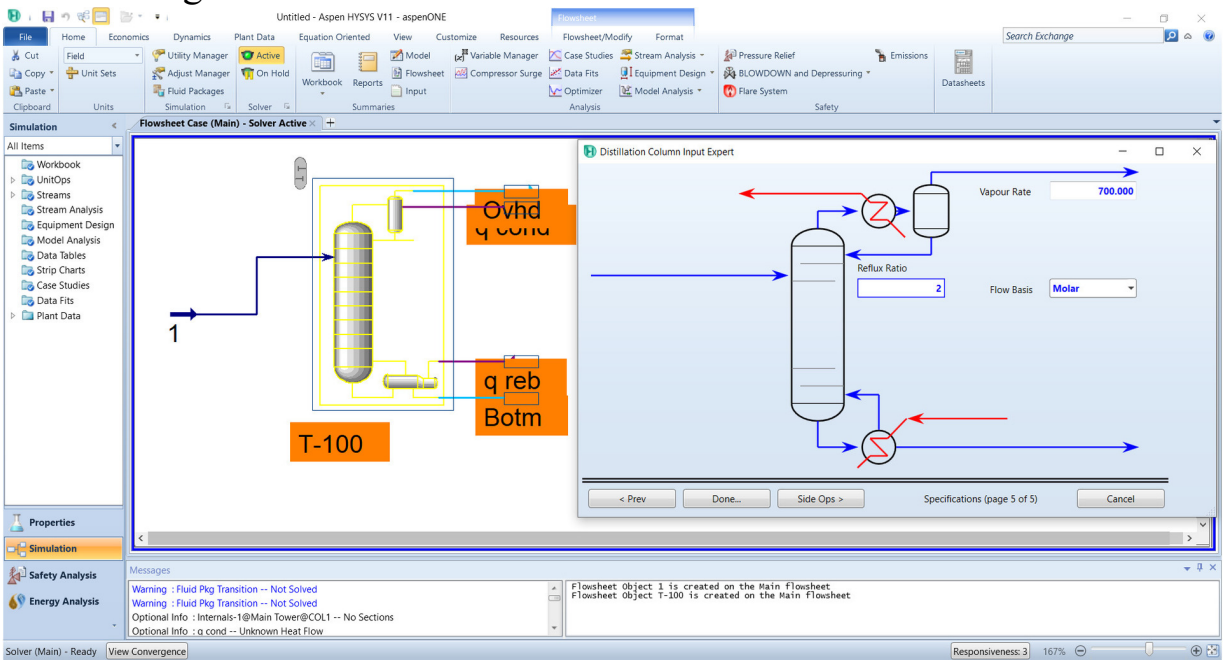
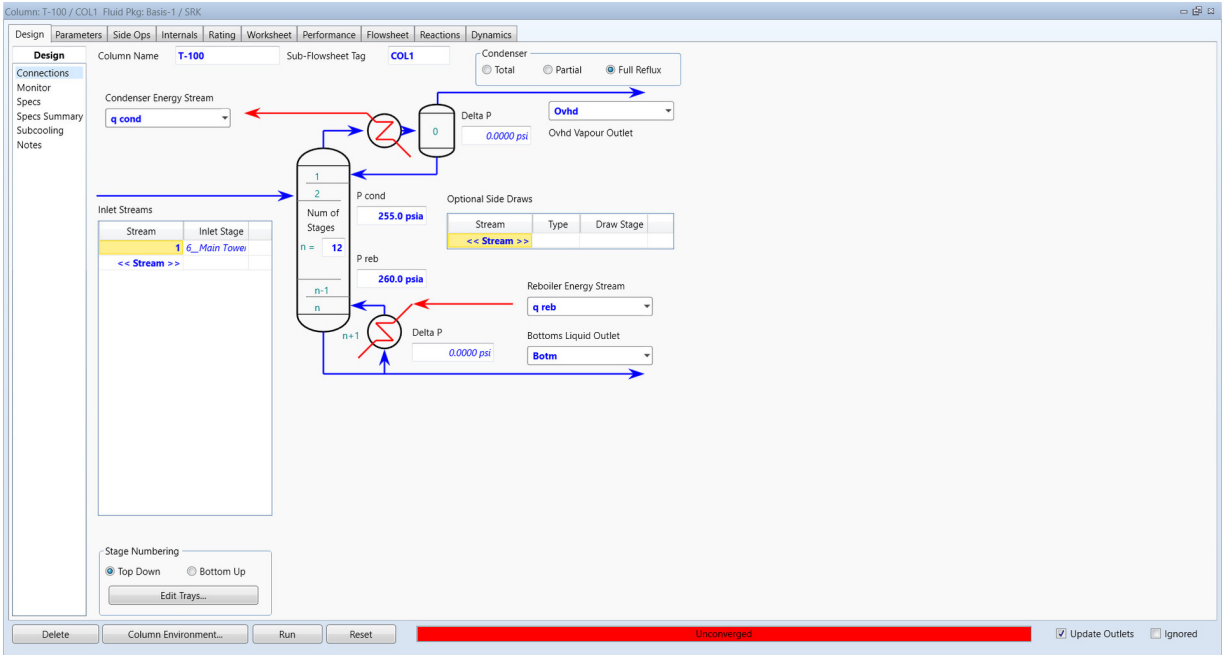
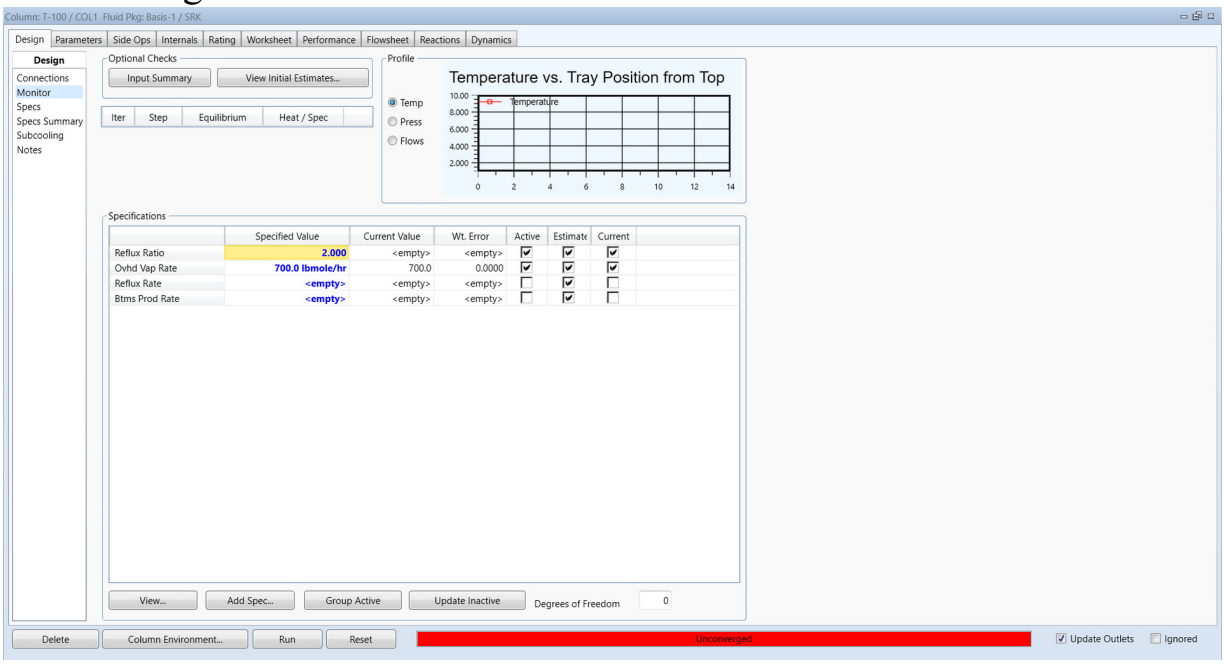


Figure 11.11: Distillation column



Click “Run”.

Figure 11.12: Distillation column



We must make 2 specifications active to make the DOF (degree of freedom) = 0 and to converge the column.
The two specifications are the target that you want to achieve from the column. Although the column is converged, it is not always practical to have vapor rate & reflux ratio specifications. These specifications can result in columns which cannot be converged or that produce product streams with undesirable properties if the column feed conditions change.

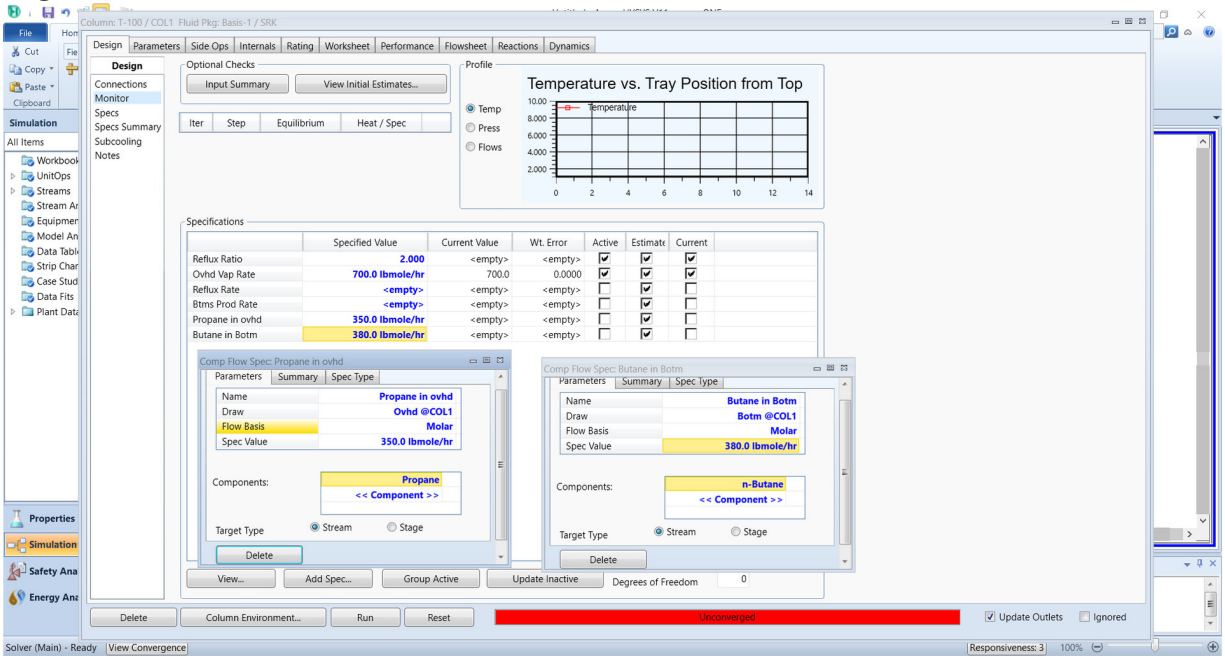
An alternative approach is to specify either component fractions or component flow rates for the column product streams.

Now we have to give the column another 2 new specifications to run with (more practical)

- Propane overhead flow of 350 lbmol/hr.
- Butane bottom flow of 380 lbmol/hr.

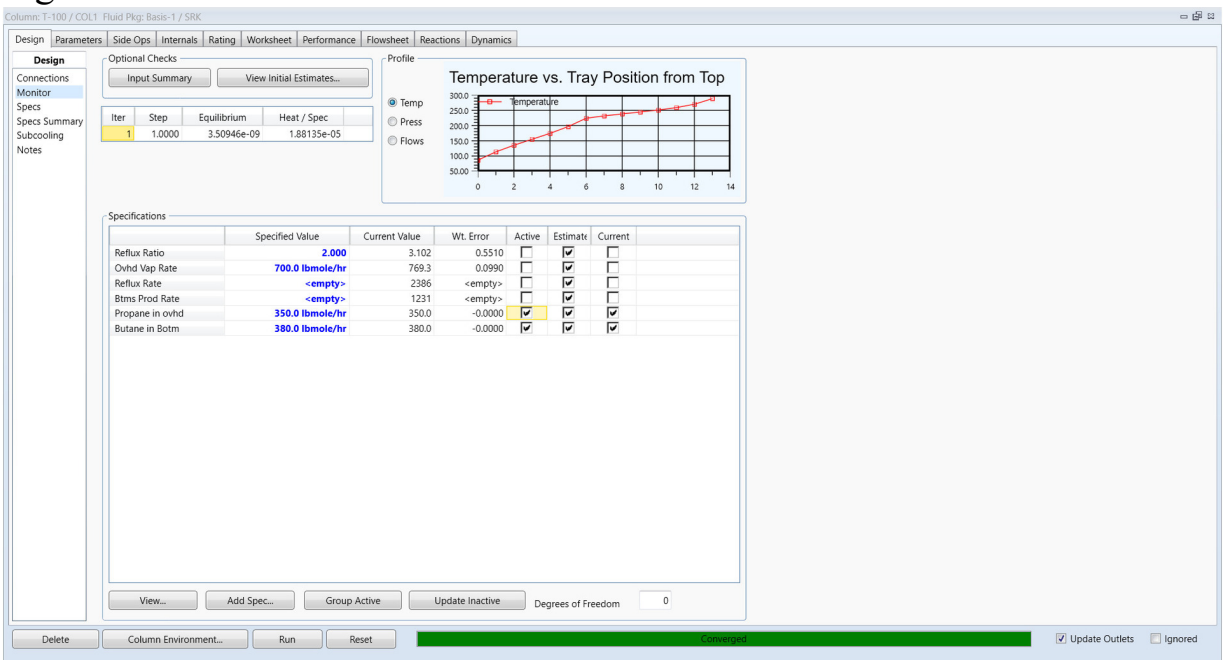
Add 2 specifications, namely Column component flow of propane in overhead product and n-Butane in bottom product.

Figure 11.13: Distillation column



Uncheck earlier active specifications and check the new specifications.

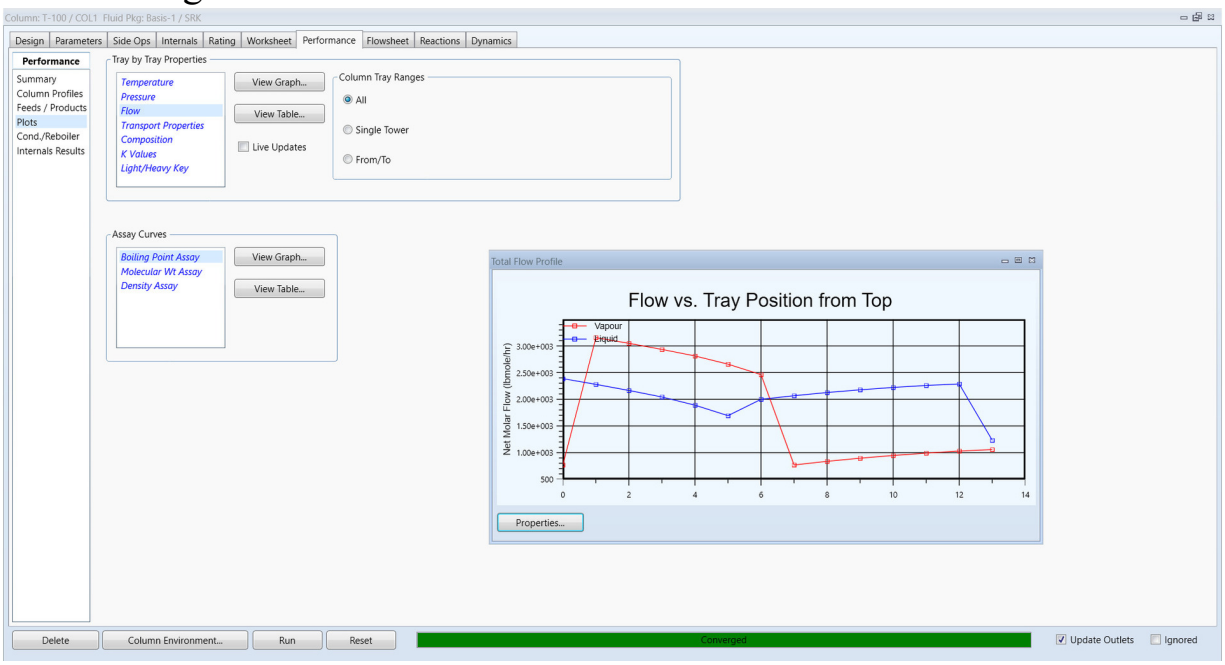
Figure 11.14: Distillation column



- Condenser Temp. : 86.4 ° F
- Reboiler Temp. : 128.8 ° C
- Reflux Ratio : 3.1

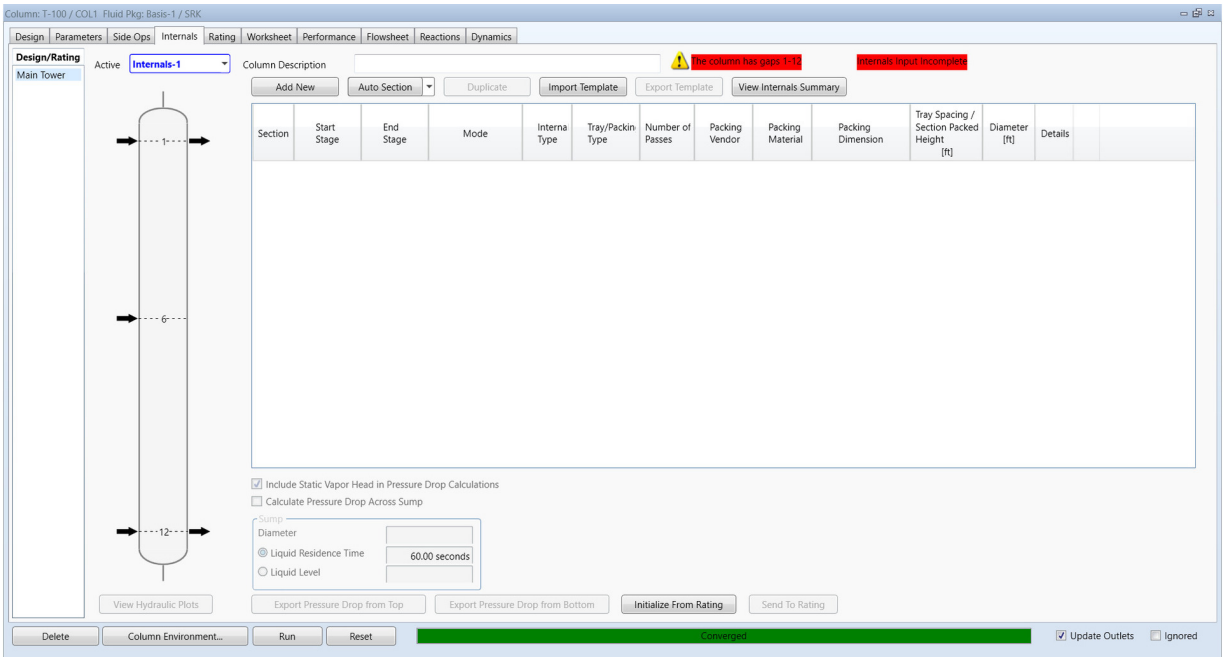
Different plots can be generated in the “Performance” tab. One such example is given below.

Figure 11.15: Distillation column



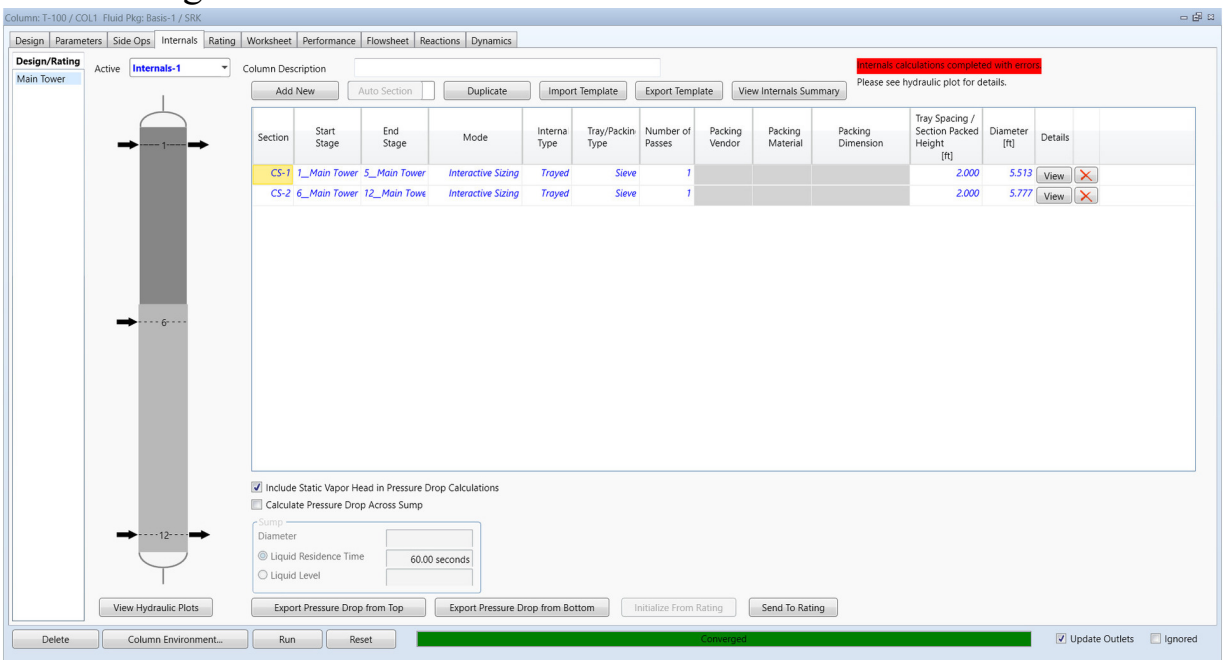
Let’s now design this distillation column for this condition. Select “Internal” tab.

Figure 11.16: Distillation column



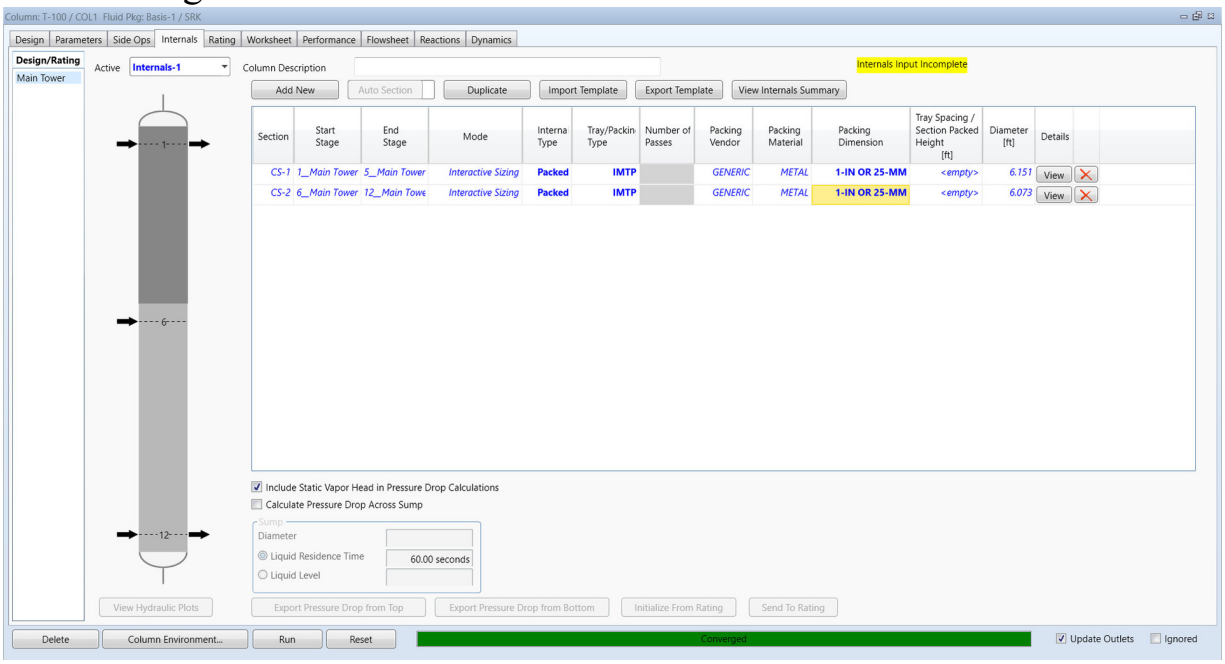
Click on “Auto Section”. It will divide 12 tray columns into 2 sections (1 to 5 trays and 6 to 12 trays).

Figure 11.17: Distillation column



Change internal type as “Packed”, Tray packing type as “IMTP” and Packing dimension as “1 inch”. Click on “View”.

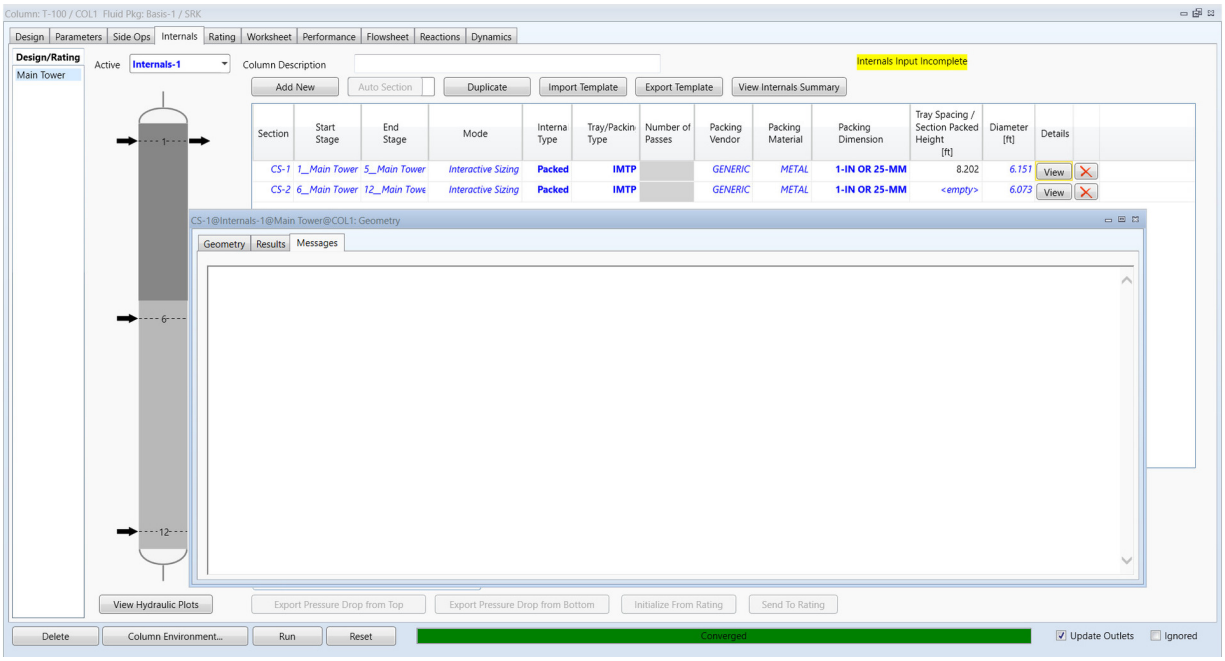
Figure 11.18: Distillation column



Select “Packed Height per Stage (HETP). HETP value of 1” IMTP packing is to be taken from the vendor. Assuming HETP value as 0.5 m or 1.64 ft, the same value is provided. Accordingly, total bed height will be calculated for section 1 (from 1st to 5th tray region).

Check the “Messages” tab. All the warnings/errors will be listed here.

Figure 11.19: Distillation column



Carry out the same procedure for CS-2.

Figure 11.20: Distillation column

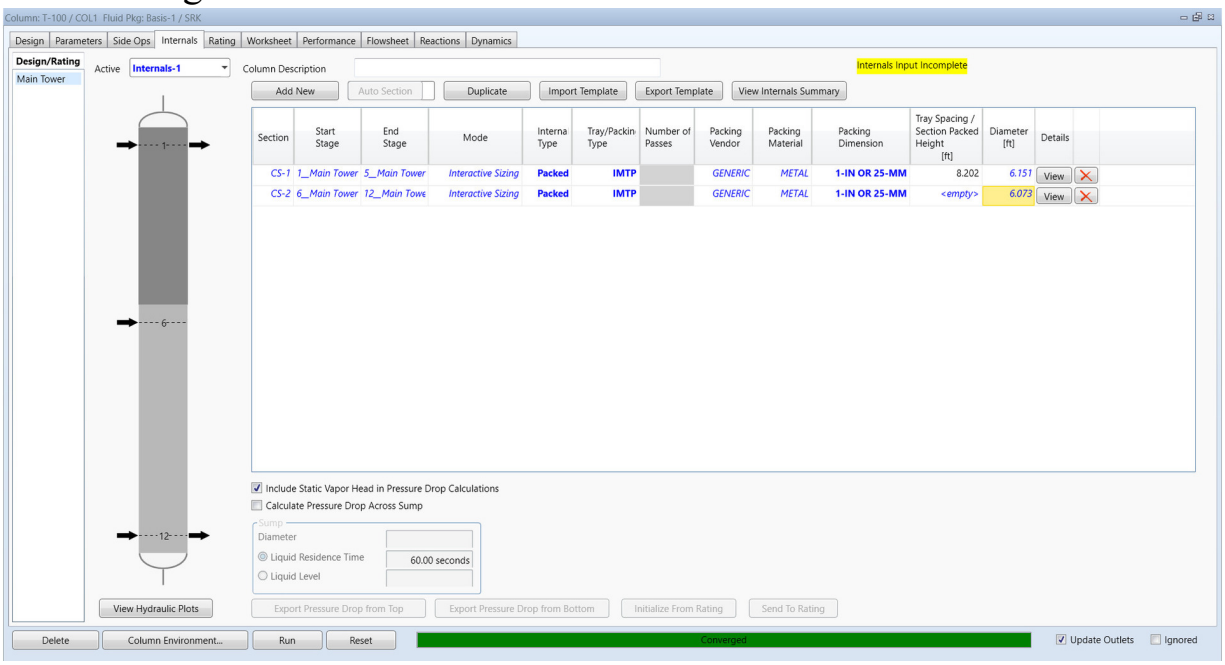


Figure 11.21: Distillation column

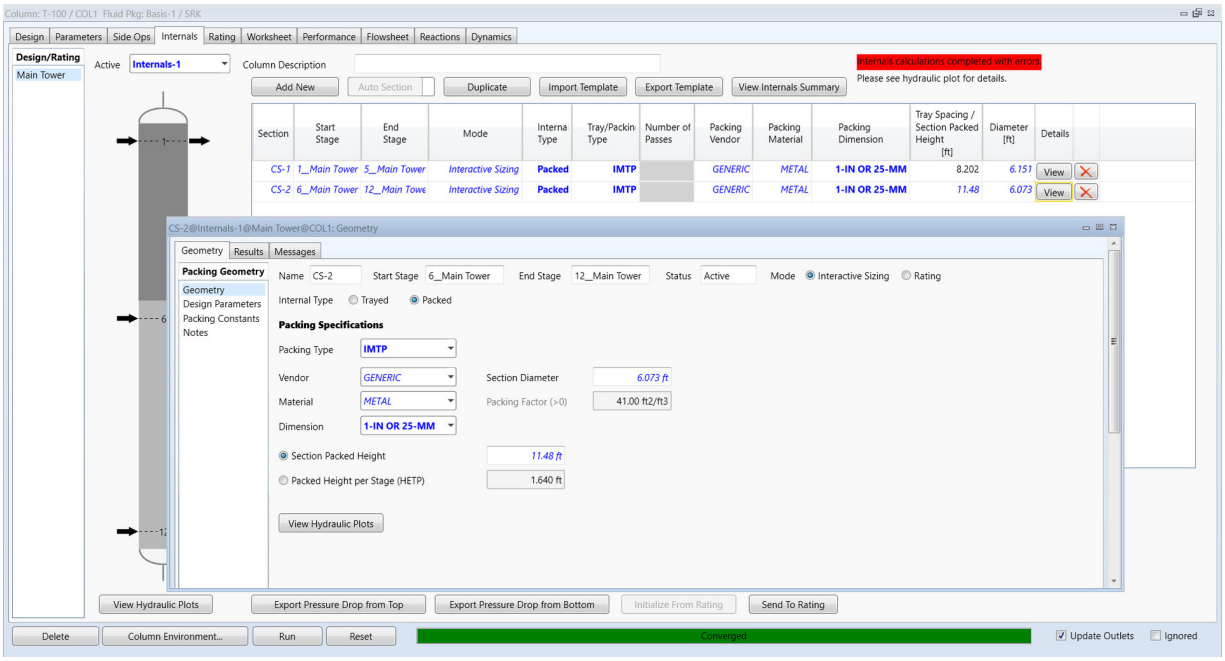
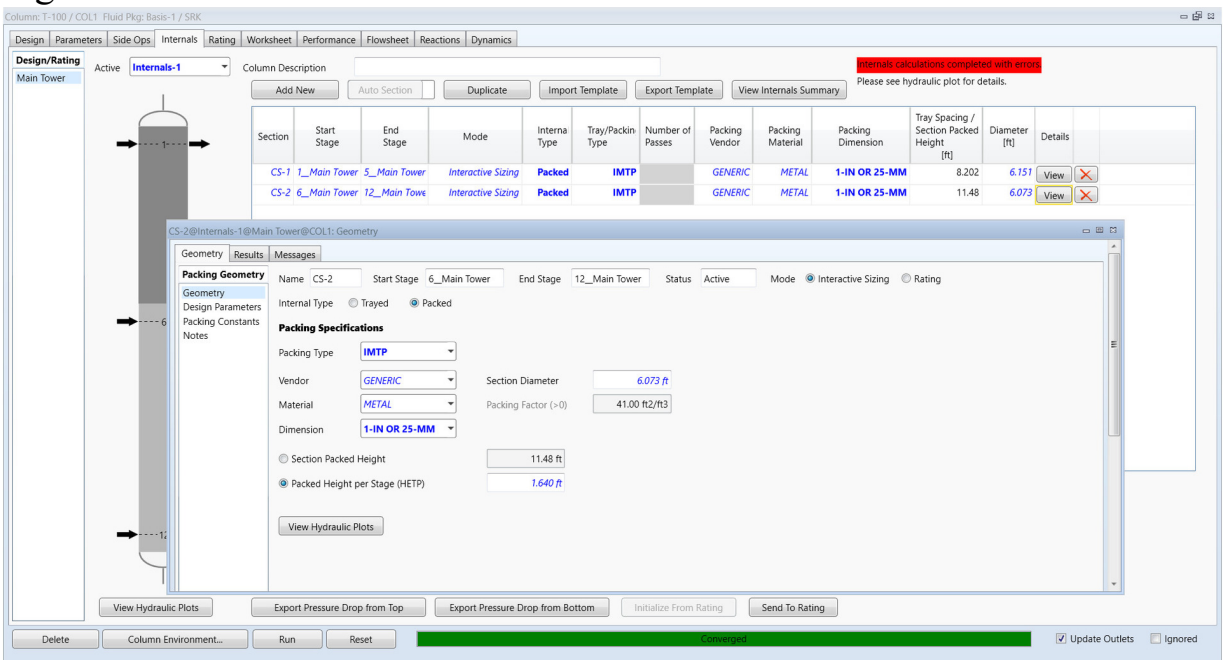
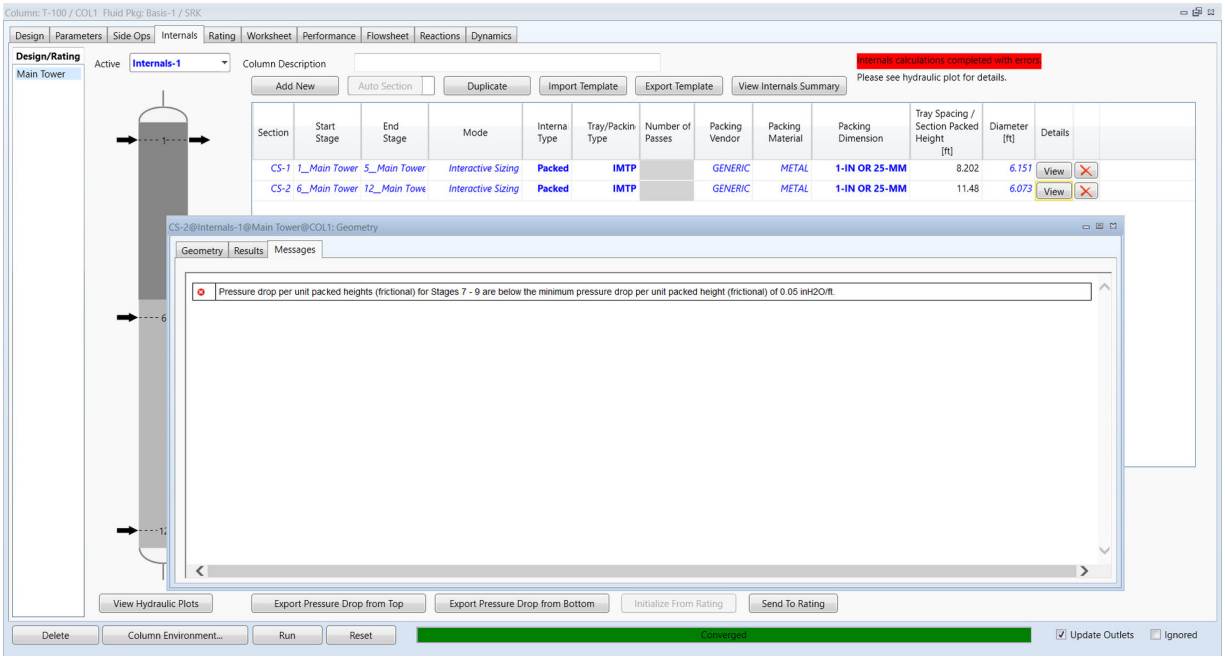

Column: T-100 / COL1 Fluid Pkg: Basis-1 / SRK
Design
Parameters
Side Ops
Internals
Rating
Worksheet
Performance
Flowsheet
Reactions
Dynamics
Design/Rating
Main Tower
Active
Internals-1
Column Description
Add New
Auto Section
Duplicate
Import Template
Export Template
View Internals Summary
Internals calculations completed with errors
Please see hydraulic plot for details.
Section
Start Stage
End Stage
Mode
Interna Type
Tray/Packin Type
Number of Passes
Packing Vendor
Packing Material
Packing Dimension
Tray Spacing / Section Packed Height [ft]
Diameter [ft]
Details
CS-1 1_Main Tower 5_Main Tower Interactive Sizing Packed IMTP GENERIC METAL 1-IN OR 25-MM 8.202 6.151 View
CS-2 6_Main Tower 12_Main Towe Interactive Sizing Packed IMTP GENERIC METAL 1-IN OR 25-MM 11.48 6.073 View
CS-2@Internals-1@Main Tower@COL1: Geometry
Geometry
Results
Messages
Packing Geometry
Geometry
Design Parameters
Packing Constants
Notes
Name CS-2
Start Stage 6_Main Tower
End Stage 12_Main Tower
Status Active
Mode Interactive Sizing Rating
Internal Type Trayed Packed
Packing Specifications
Packing Type IMTP
Vendor GENERIC
Material METAL
Dimension 1-IN OR 25-MM
Section Diameter 6.073 ft
Packing Factor (>0) 41.00 ft2/ft3
Section Packed Height 11.48 ft
Packed Height per Stage (HETP) 1.640 ft
View Hydraulic Plots
View Hydraulic Plots
Export Pressure Drop from Top
Export Pressure Drop from Bottom
Initialize From Rating
Send To Rating
Delete
Column Environment...
Run
Reset
Converged
Update Outlets
Ignored

Figure 11.22: Distillation column



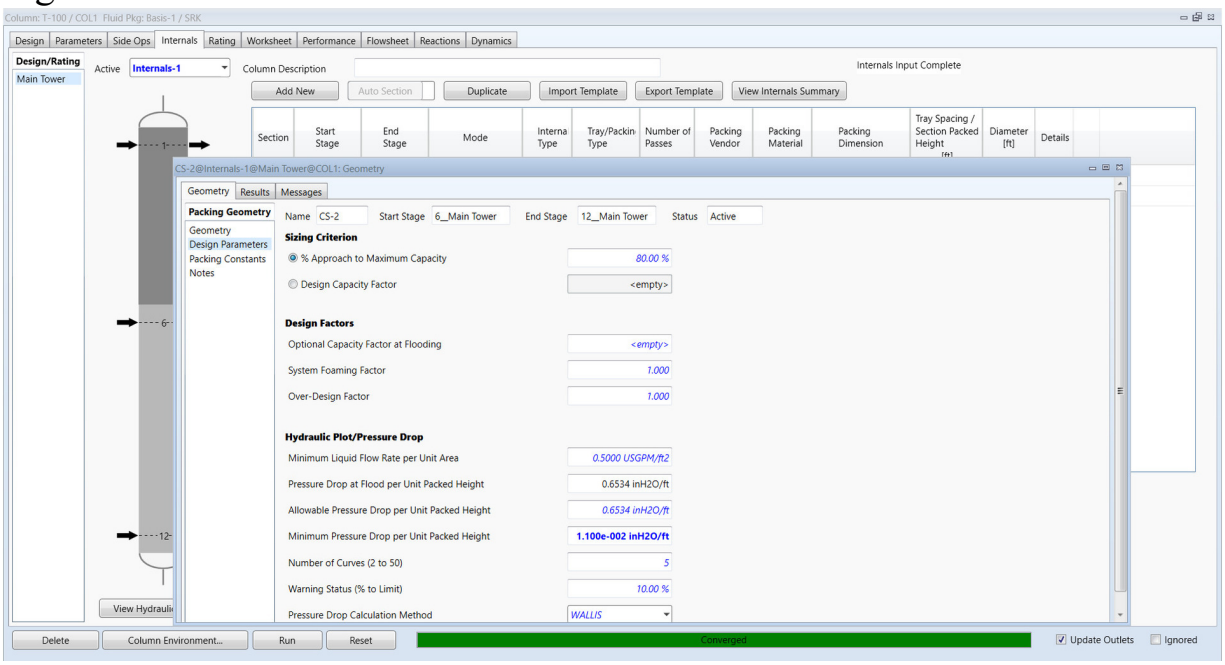
The error message states that the pressure drop calculated per bed height is less than the default value.

Figure 11.23: Distillation column



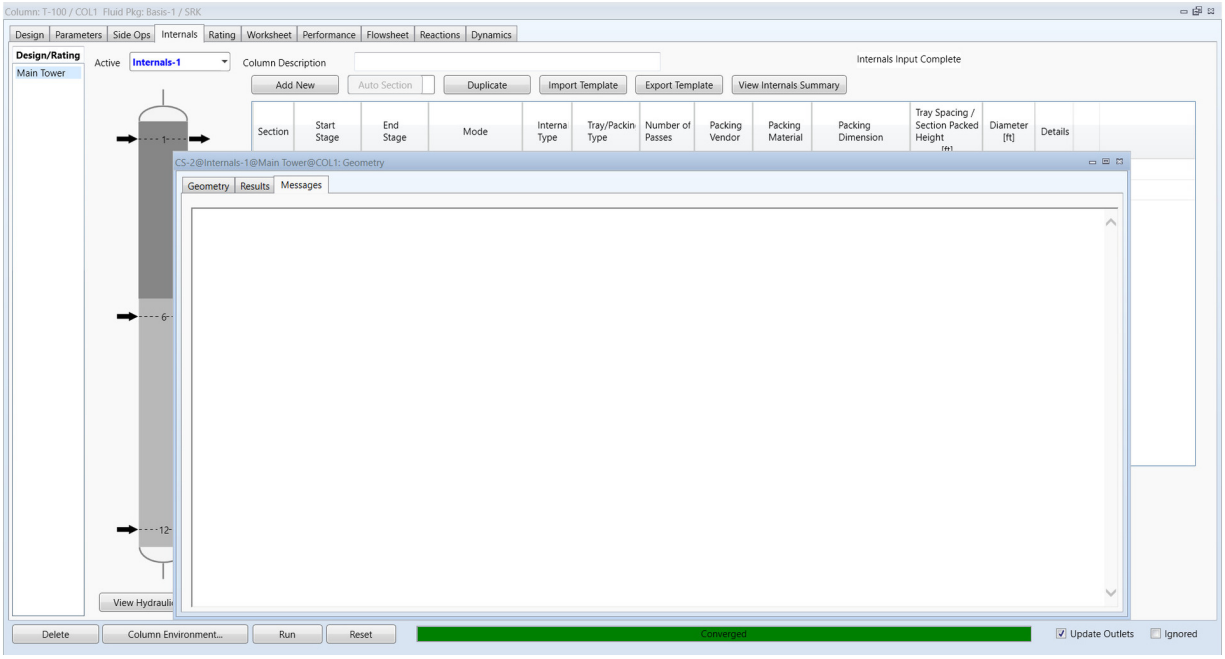
The default value can be changed with the consultation with the vendor. Assuming the vendor-provided value for "minimum pressure drop per unit packed height" is 0.01 inch H_2 O/ft.

Figure 11.24: Distillation column



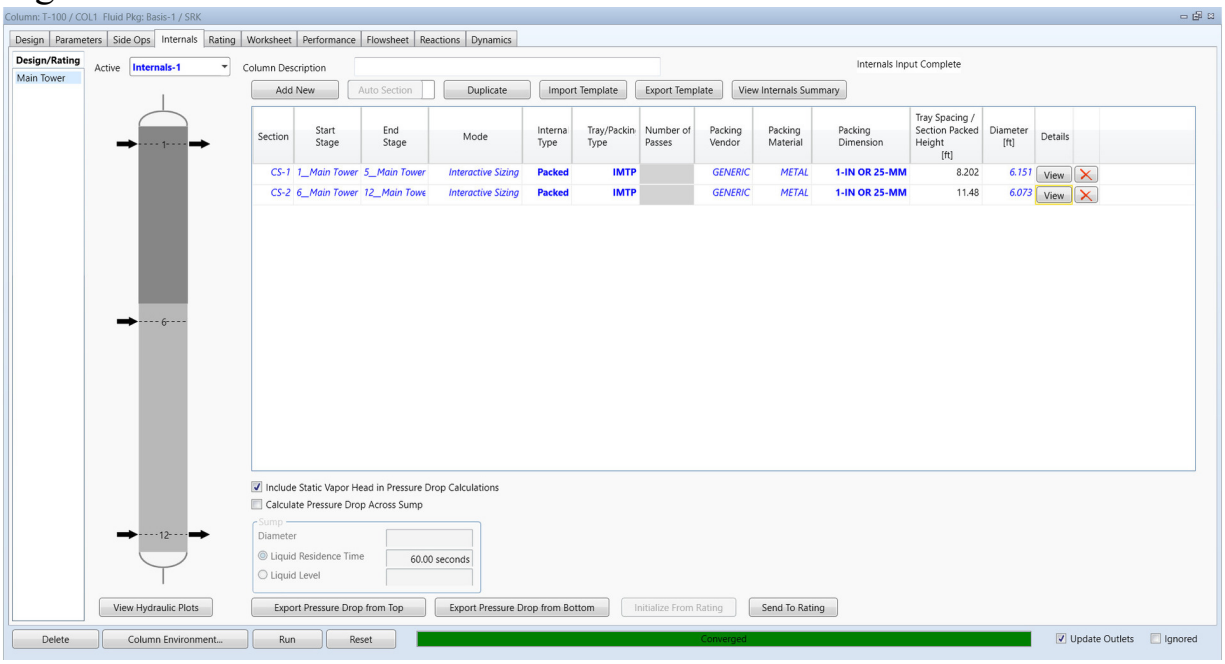
This change has resolved the error.

Figure 11.25: Distillation column



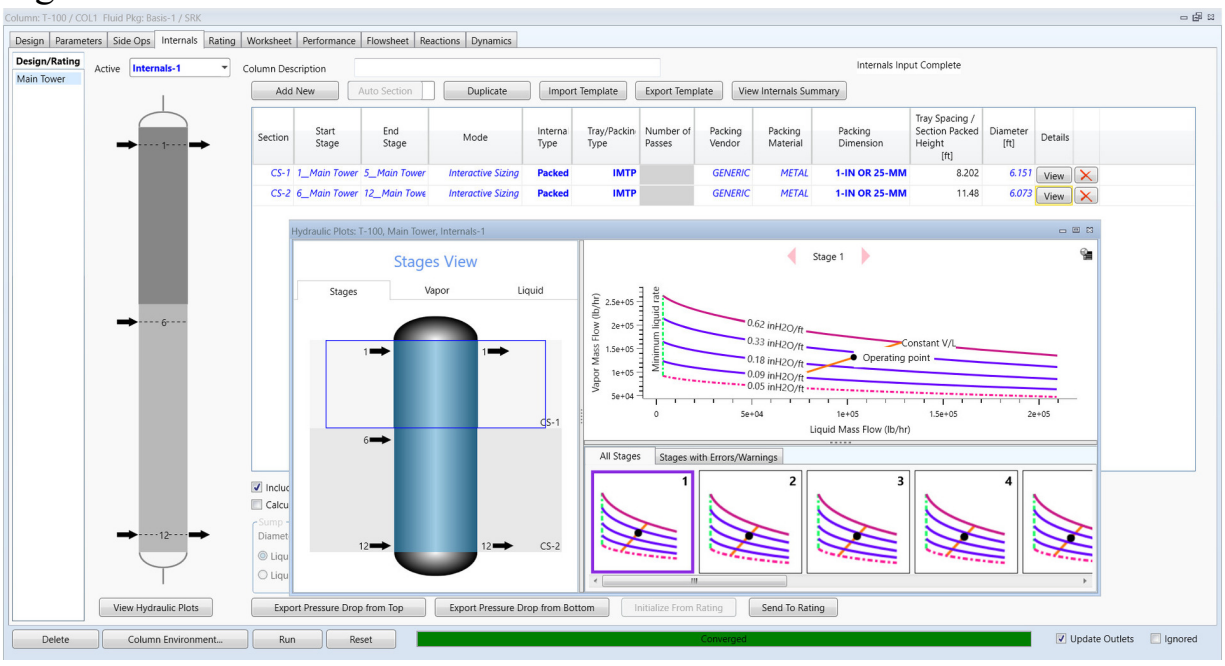
Column diameter for CS-1 and CS-2 are 6.149 ft and 6.071 ft respectively.

Figure 11.26: Distillation column



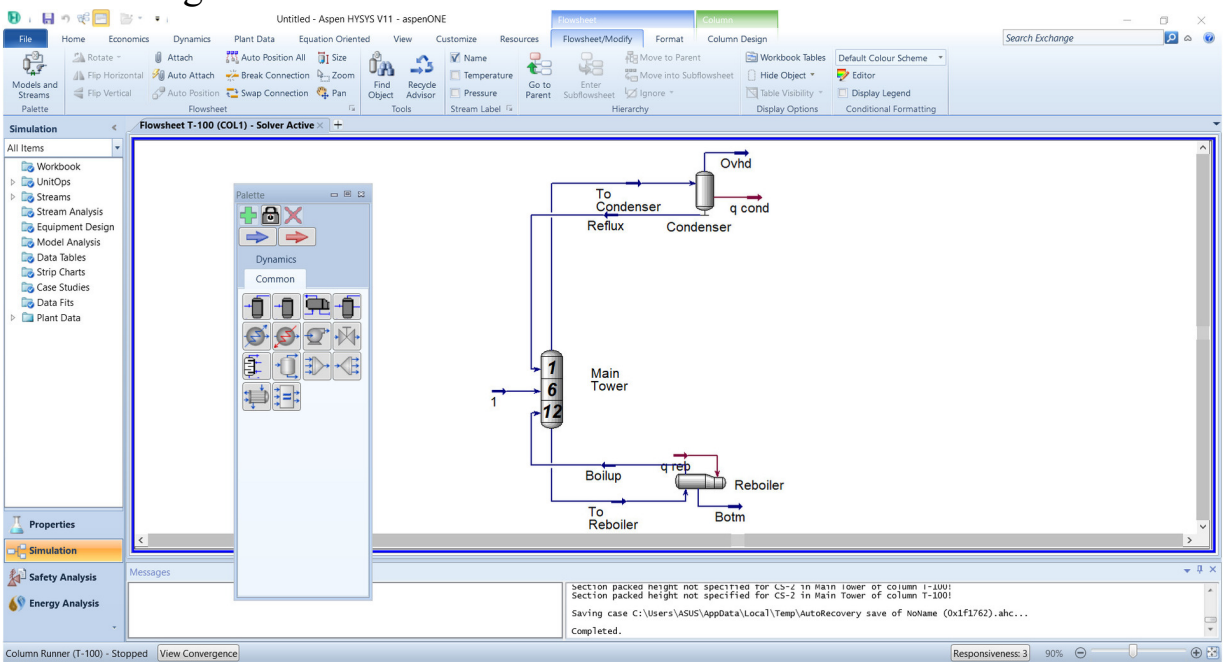
Different hydraulic plots can also be generated in Aspen HYSYS.

Figure 11.27: Distillation column



Clicking “Column Environment” will take you to column flowsheet with internal streams.

Figure 11.28: Distillation column



Exercises

a. Mr. Nitin needs to simulate a distillation column which has the same process parameters as given in the solved example. He has, however, a tray column with 36 actual trays with 50% efficiency, available in his plant. What do you think would be reflux ratio, condenser and reboiler duty in this case? Without carrying out any simulation, is it possible for you to answer that question qualitatively?

b. Mr. Ranjit is Mr. Nitin's colleague and has been tasked to convert the available tray column into a packed column. What are the different options he has with packings? Can the same column be used?

c. Mr. Amit wants to run the solved example with same column and process parameters, but with 50% more feed flow rate. Do you expect to get

50% more product yield? If not, what other measures should be taken by him?

d. Mr. Himanshu is required to carry out a comparative study of solved example w.r.to the effect of reflux ratio on reboiler duty and condenser duty. Can you help him with that assignment?
e. Mr. Rohan wants to convert beer into whiskey. Can you make a problem statement and solve it so that Mr. Rohan can use it for reference?
f. Ms. Madhugandha needs to simulation the ammonia absorption column. Can you make a problem statement and solve it so that Ms. Madhugandha can use it for reference?
g. Mr. Adeeb wants to learn about the gas sweetening plant. Can you make a problem statement and solve it so that Mr. Adeeb can use it for reference? You may use MDEA as a solvent.
h. Mr. Payas needs to design a column to manufacture absolute alcohol. Can you make a problem statement and solve it so that Mr. Payas can use it for reference? This solution must not have any solvent.
i. Mr. Viraj needs to design a column to manufacture absolute alcohol. Can you make a problem statement and solve it so that Mr. Viraj can use it for reference? This solution should have some solvent.
j. Mr. Anil wants to learn about the BTX purification plant. Can you make a problem statement and solve it so that Mr. Anil can use it for reference?

CHAPTER 12: OPTIMIZER TOOL IN ASPEN HYSYS

In this example, a simple distillation column to separate Tetrahydrofuran (THF) from Toluene is simulated. The object of the exercise is to select the product specifications such that profit is maximized. A special tool in HYSYS, the Optimizer, will be used to find the optimum operating conditions.
HYSYS includes additional modeling and decision support tools that can be used to enhance the usability of your models. In this module, you will use the HYSYS optimization tool available in HYSYS to investigate the debottlenecking and optimization of a crude column.
Once you have completed, you will be able to;

- Use the Optimizer tool in HYSYS to optimize flow sheets
- Use the Spreadsheet to perform calculations

5000 kg/hr mixture of Tetrahydrofuran & toluene (30 mass% THF) at 20 ^{o} C and 160 kPa is to be separated by distillation to get each of them with a purity of 99.2 mass% of THF & 95 mass% of Toluene (THF is the more volatile component).
Use Wilson fluid package.
The column specifications are:

- The condenser & reboiler pressure are 113 kPa & 117 kPa.
- The condenser works on total condensation conditions.
- Number of stages = 16.
- Feed enters from the 8 th tray.

Calculate:
The reflux ratio and the distillate rate under the specified conditions.
Data:

- Feed price = 0.04 $/kg

- Pure toluene selling price = 0.12 $/kg
- Pure THF selling price = 0.35 $/ kg
- Cooling Cost = 0.43 $/ kWh
- Heating Cost = 0.75 $/kWh

Note:
Profit = (Total Toluene selling price + Total THF selling price) - (Feed cost + Heating cost + Cooling Cost)
Use a range of 0.99 to 0.999 for the THF limit & 0.9 to 0.99 for the toluene.
Add components and select the fluid package.

Figure 12.1: Components

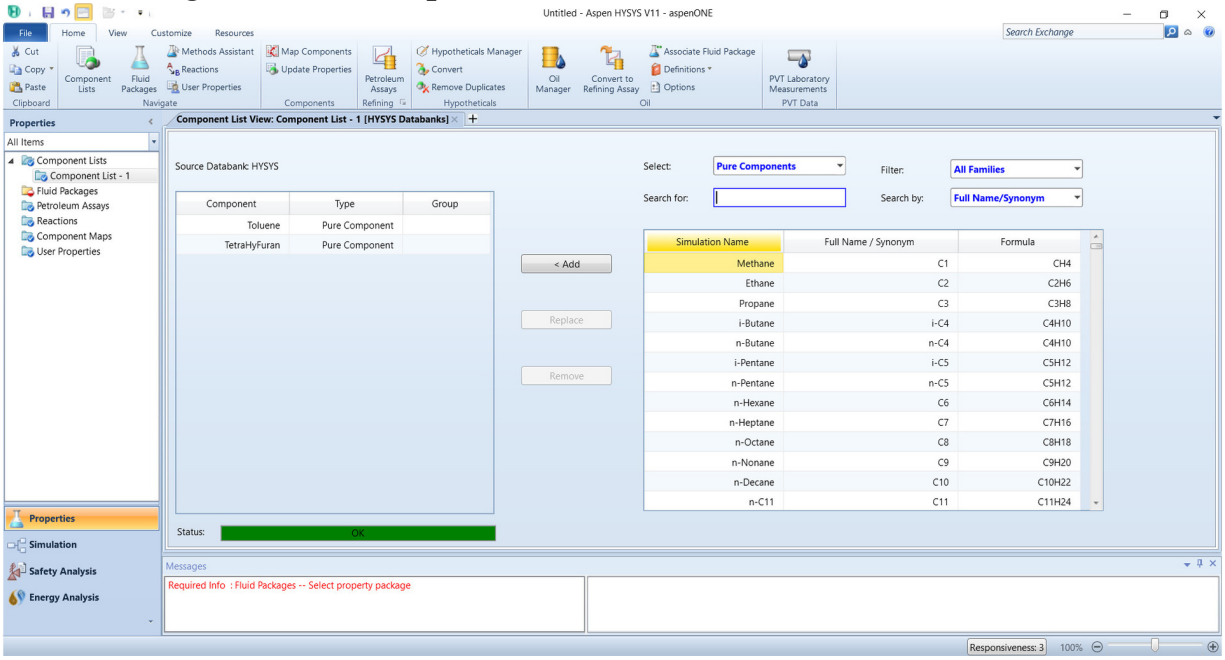


Figure 12.2: Fluid package

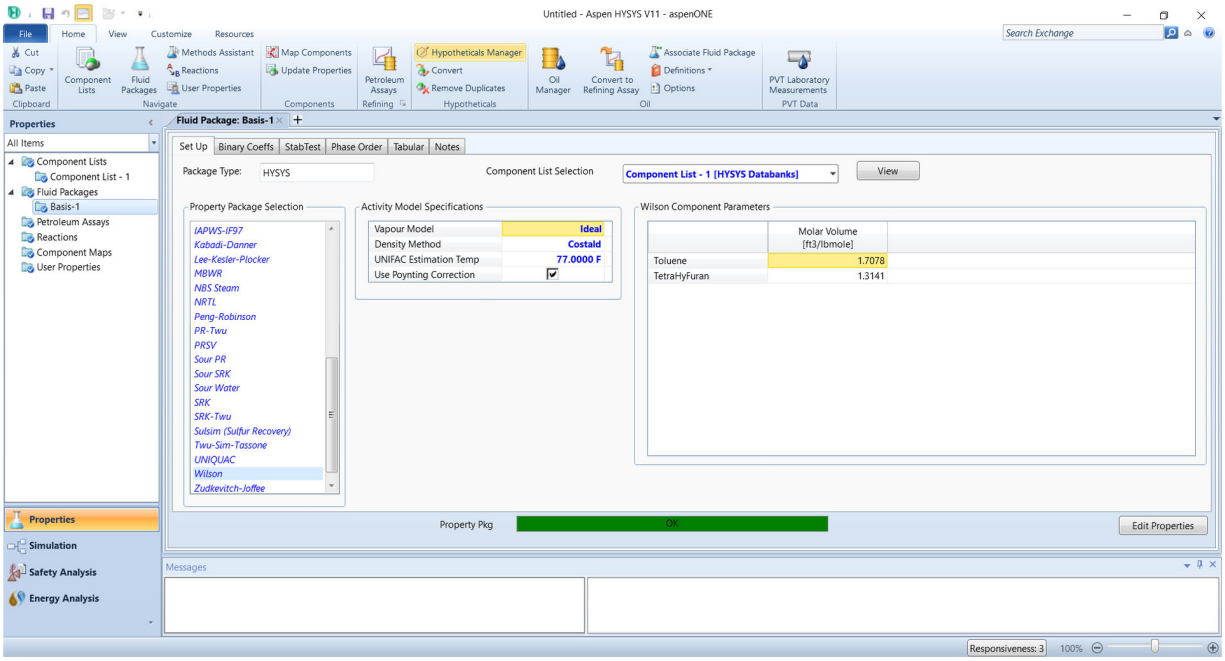
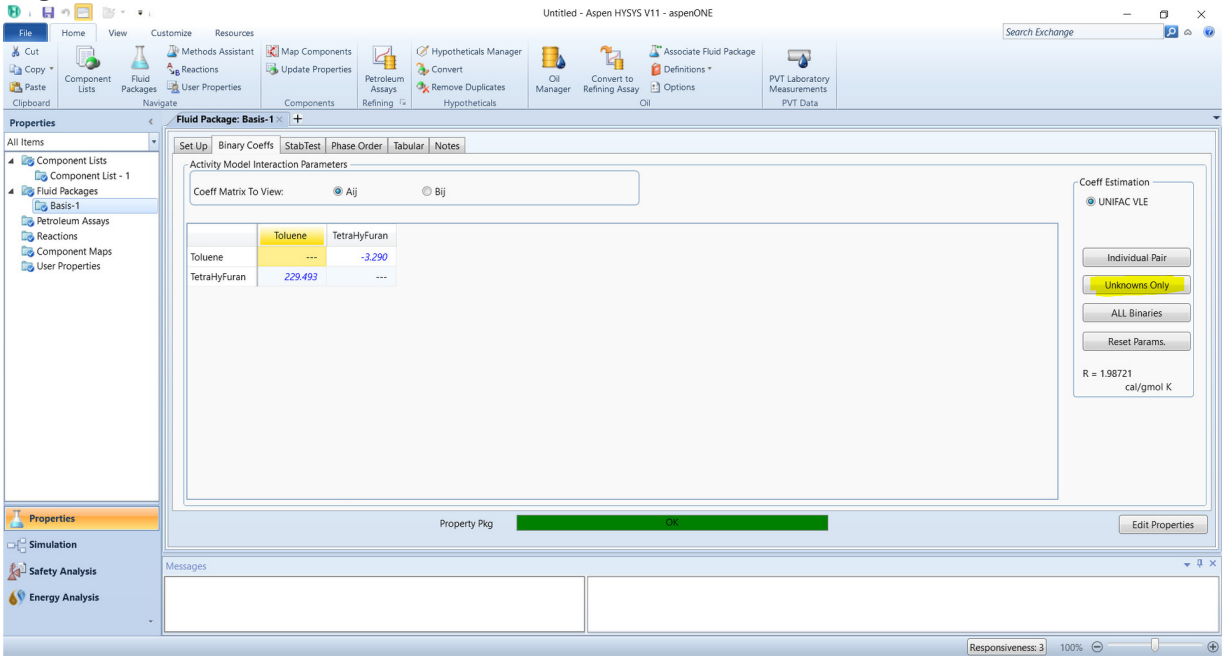
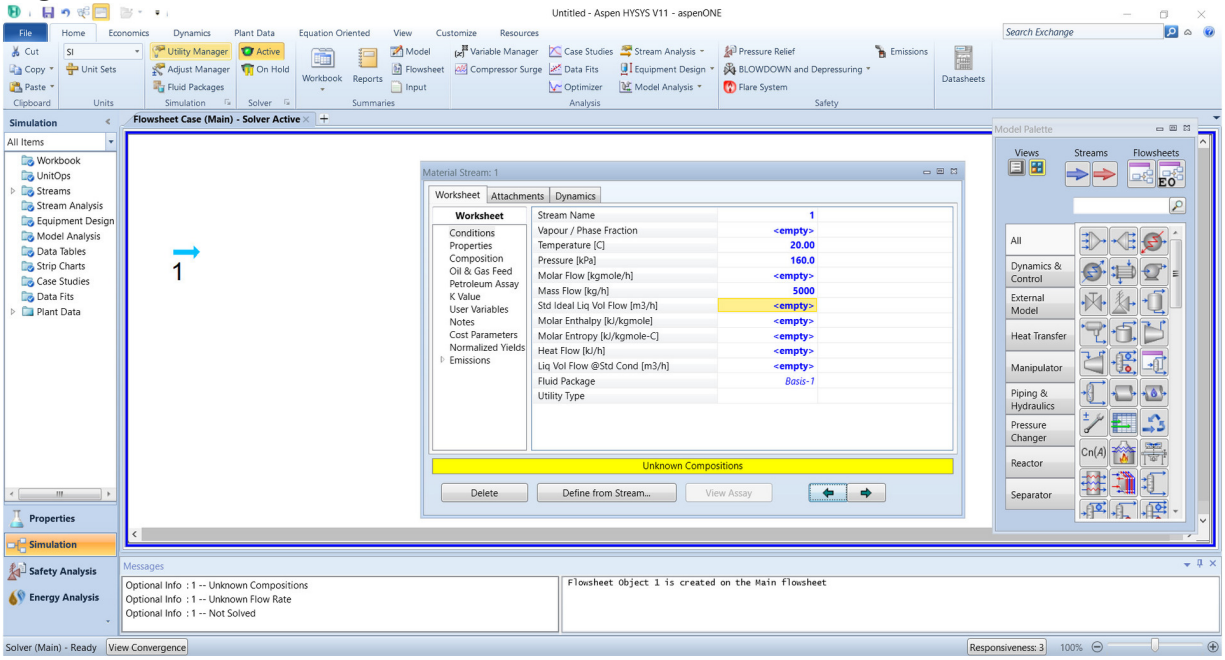

Untitled - Aspen HYSYS V11 - aspenONE
File
Home
View
Customize
Resources
Search Exchange
Cut
Copy
Paste
Clipboard
Component Lists
Fluid Packages
Methods Assistant
Reactions
User Properties
Navigate
Map Components
Update Properties
Components
Petroleum Assays
Refining
Hypotheticals Manager
Convert
Remove Duplicates
Hypotheticals
Oil Manager
Convert to Refining Assay
Associate Fluid Package
Definitions
Options
Oil
PVT Laboratory Measurements
PVT Data
Properties
All Items
Component Lists
Component List - 1
Fluid Packages
Basis-1
Petroleum Assays
Reactions
Component Maps
User Properties
Fluid Package: Basis-1
Set Up
Binary Coeffs
StabTest
Phase Order
Tabular
Notes
Package Type:
HYSYS
Component List Selection
Component List - 1 [HYSYS Databanks]
View
Property Package Selection
IAPWS-IF97
Kabadi-Danner
Lee-Kesler-Plocker
MBWR
NBS Steam
NRTL
Peng-Robinson
PR-Twu
PRSV
Sour PR
Sour SRK
Sour Water
SRK
SRK-Twu
Sulsim (Sulfur Recovery)
Twu-Sim-Tassone
UNIQUAC
Wilson
Zudkevitch-Joffee
Activity Model Specifications
Vapour Model
Ideal
Density Method
Costald
UNIFAC Estimation Temp
77.0000 F
Use Poynting Correction
Wilson Component Parameters
Molar Volume [ft3/lbmole]
Toluene
1.7078
TetraHyFuran
1.3141
Property Pkg
OK
Edit Properties
Properties
Simulation
Safety Analysis
Energy Analysis
Messages
Responsiveness: 3
100%

Figure 12.3: BIP



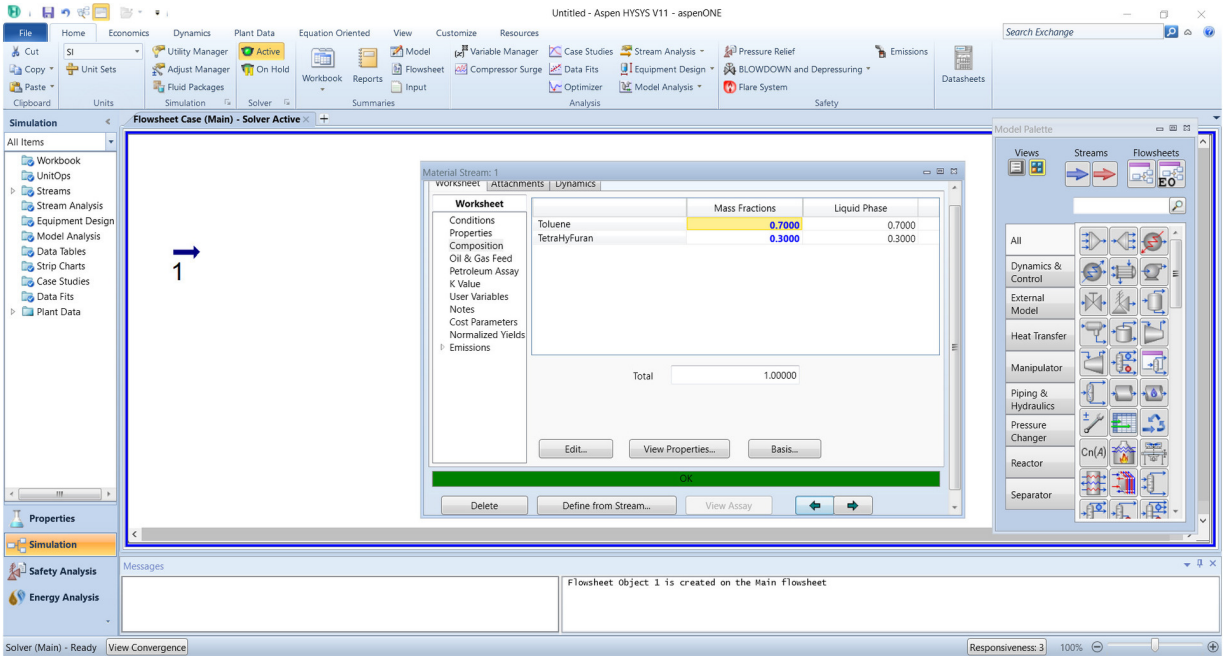
Now you can start drawing the flow sheet for the process by clicking the Simulation button. Add a material stream to define the composition and the conditions of the feed stream.

Figure 12.4: Material stream



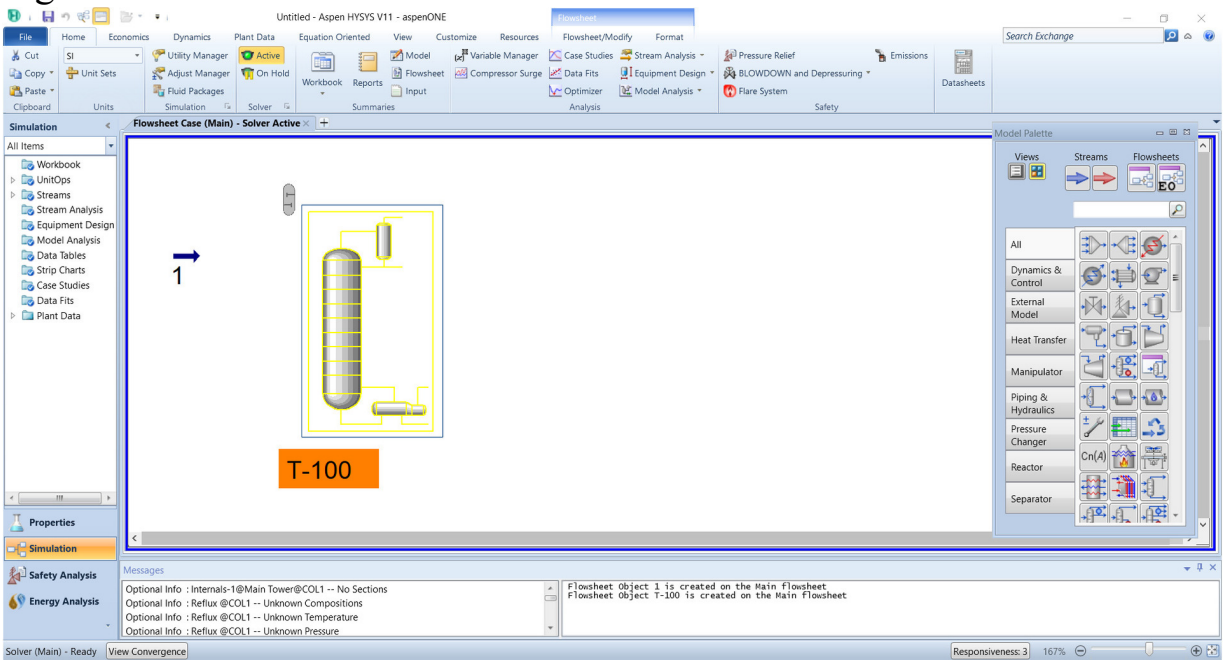
Click on “Composition” >> “Basis”. Change Basis to “Mass Fraction”.

Figure 12.5: Material stream



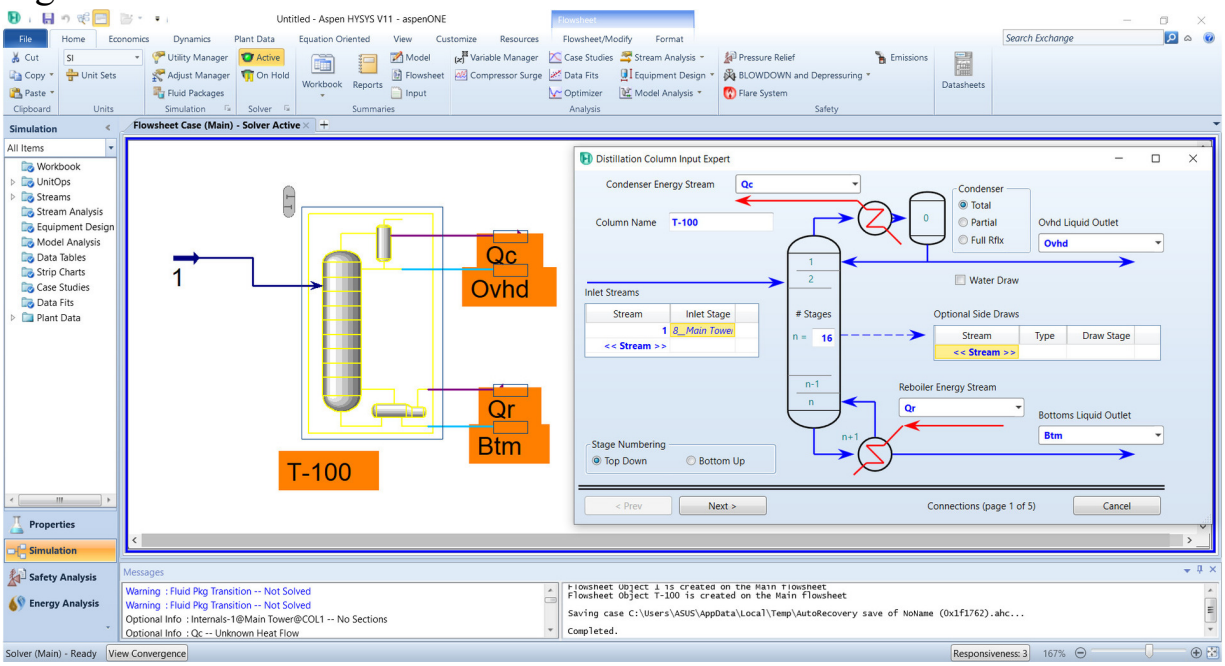
Click Ok.
Add a “Distillation column”.

Figure 12.6: Distillation column



Double click on Column (T-100) and provide details.

Figure 12.7: Distillation column



Click “Next”.

Figure 12.8: Distillation column

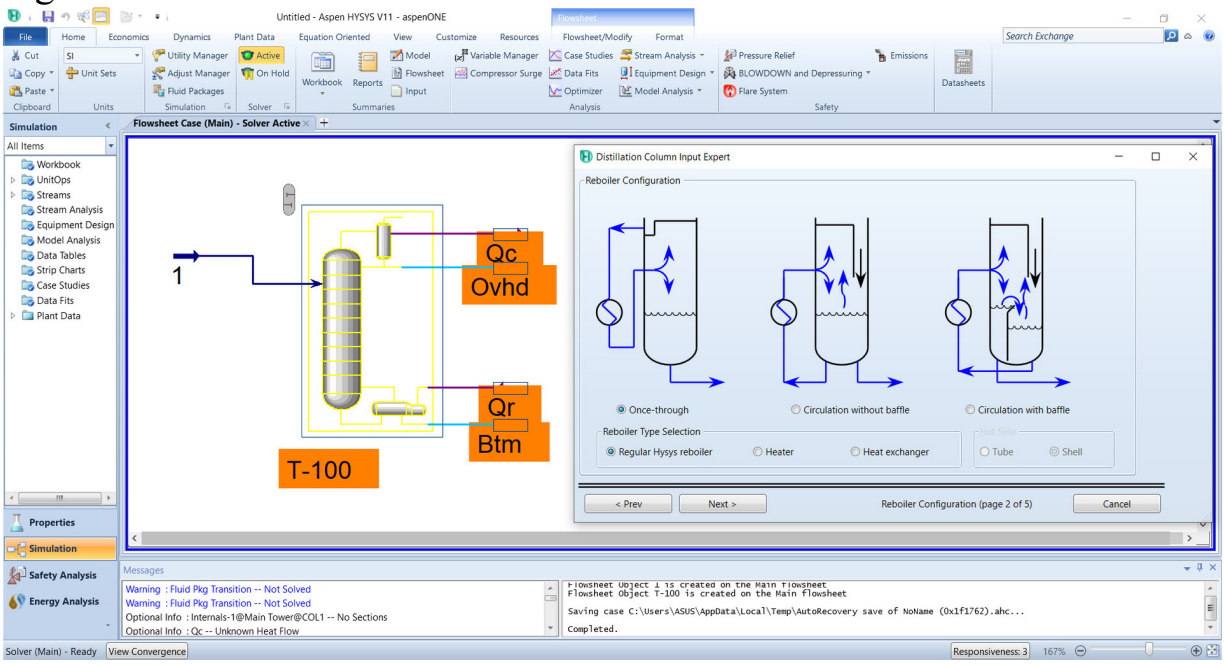


Figure 12.9: Distillation column

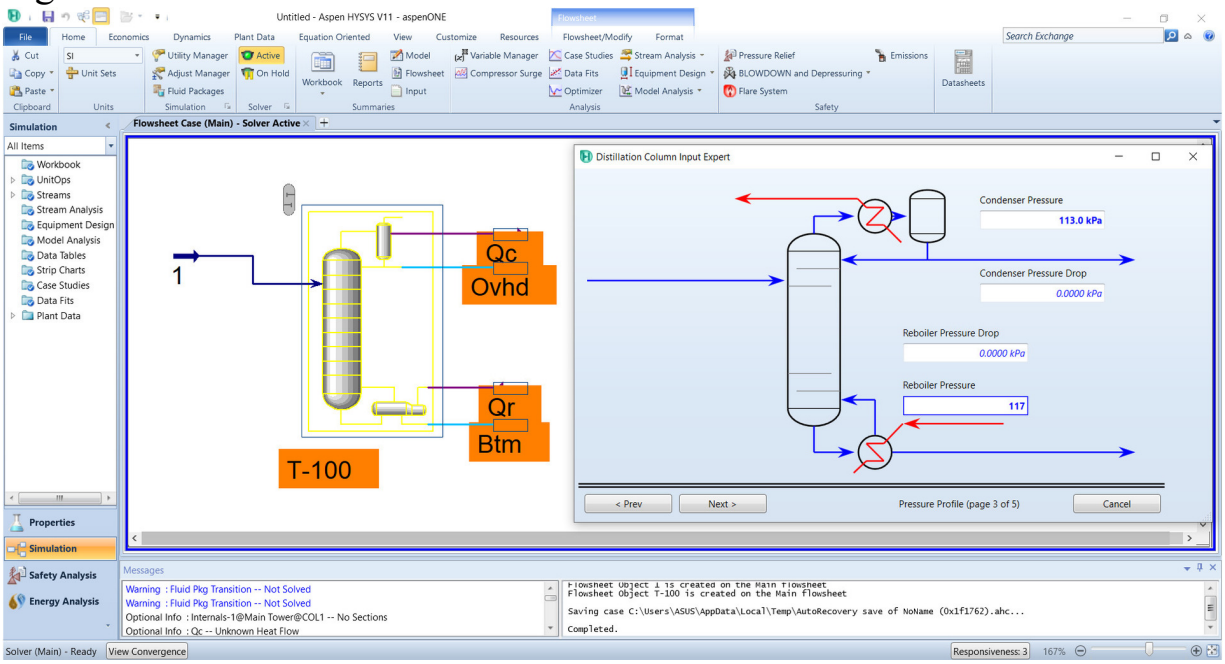


Figure 12.10: Distillation column

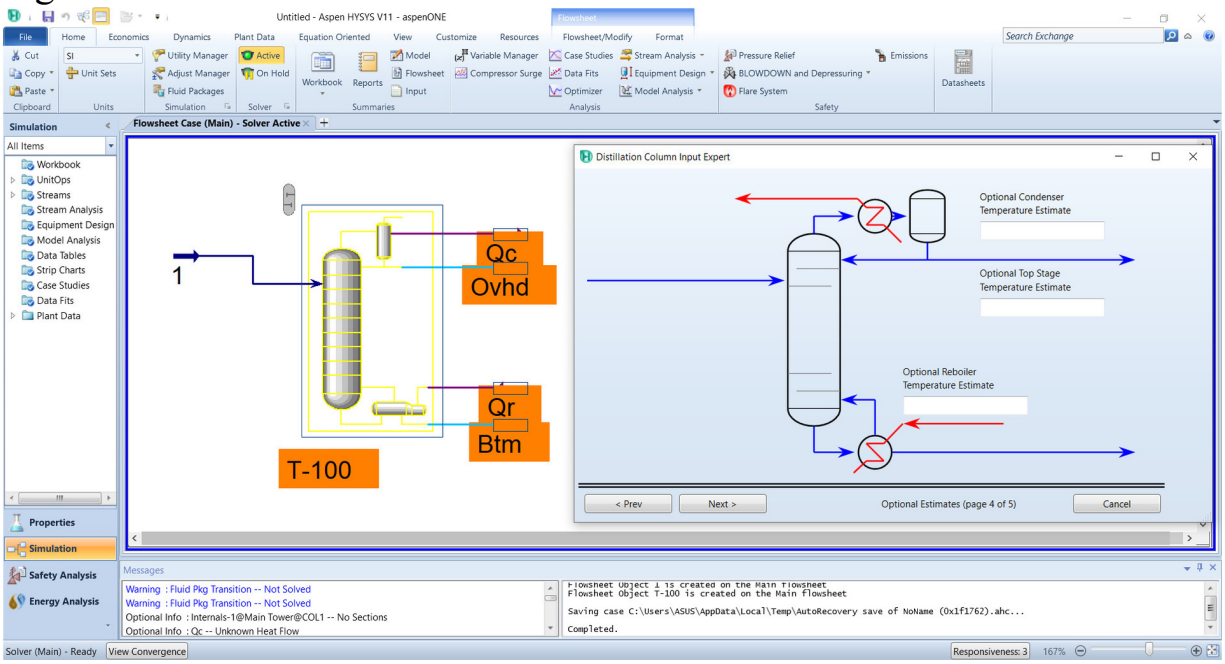
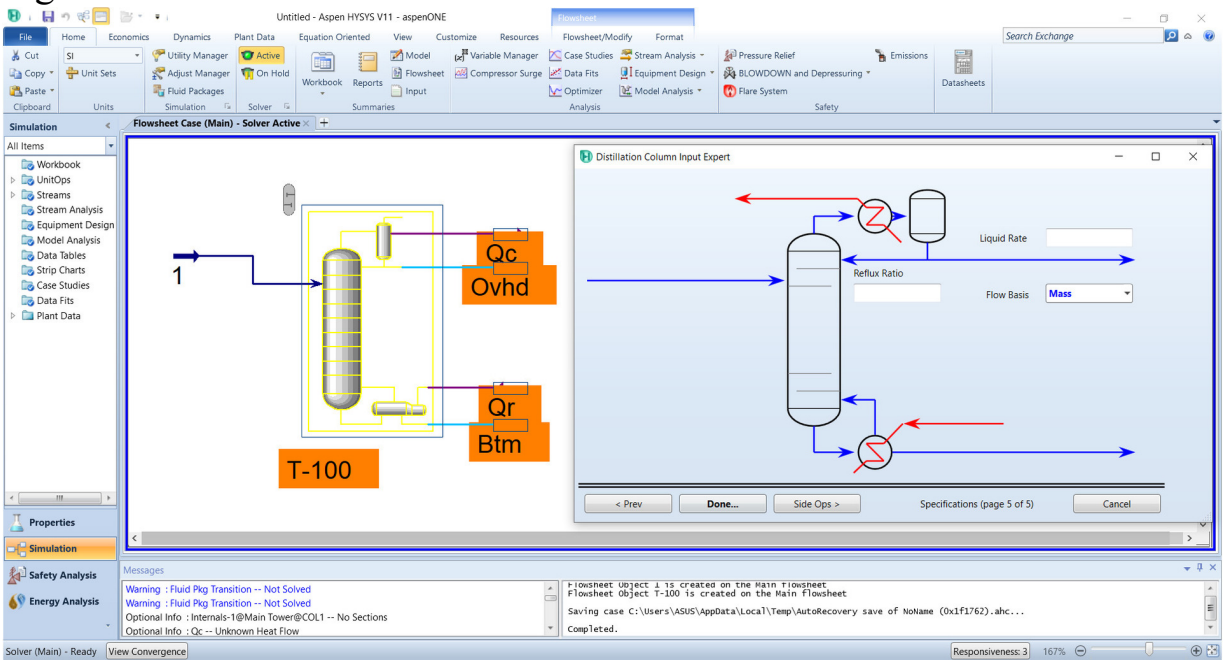
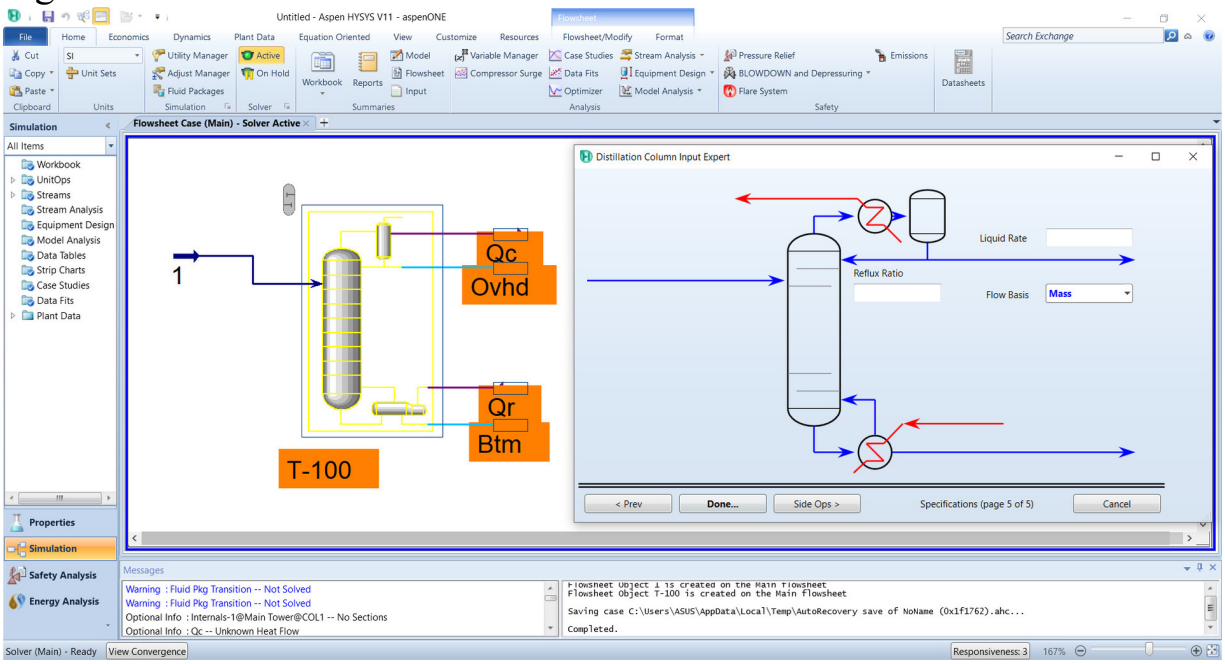


Figure 12.11: Distillation column



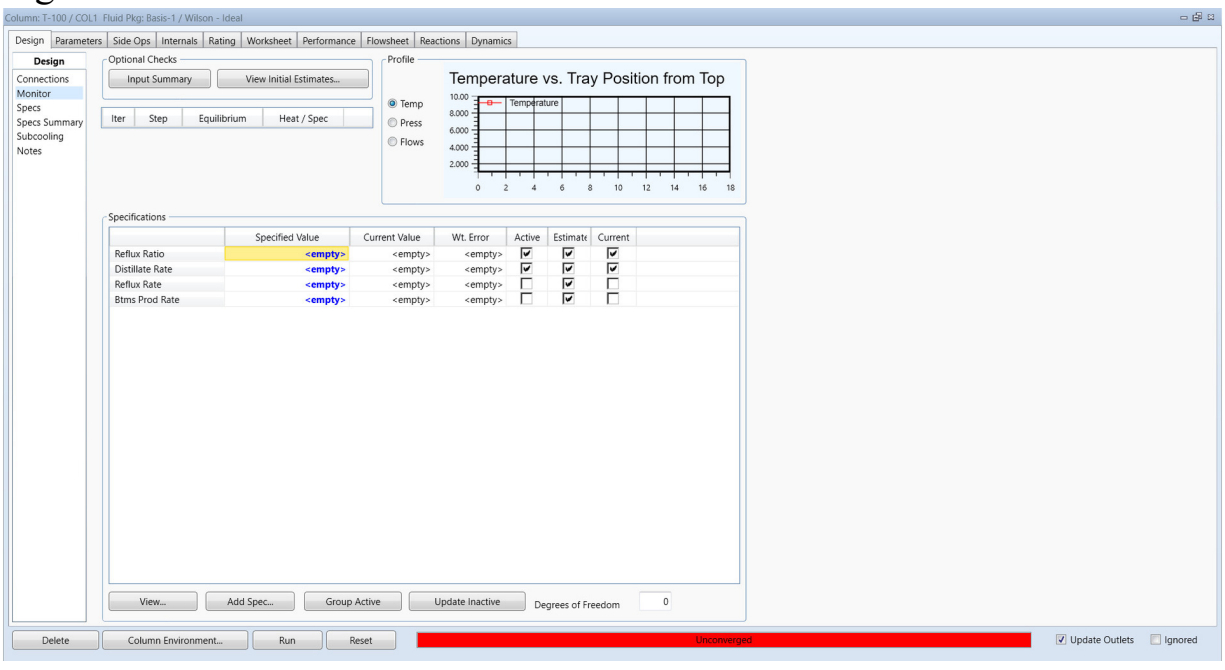
Click “Done”.

Figure 12.12: Distillation column



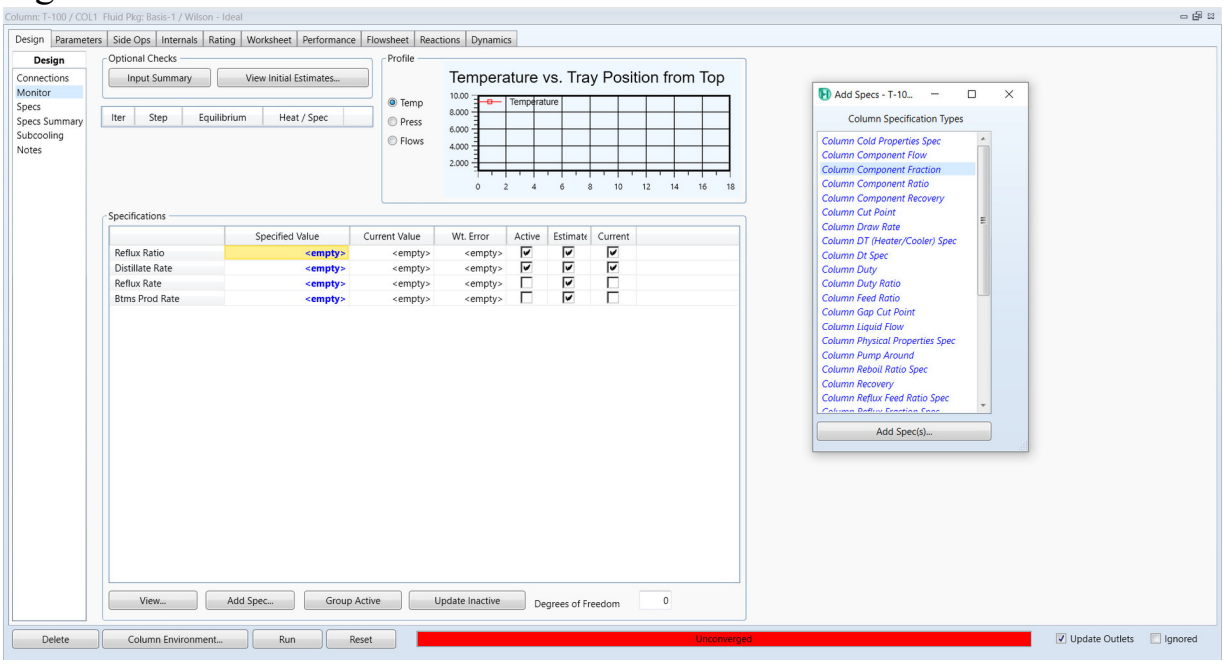
Click “Monitor”.

Figure 12.13: Distillation column



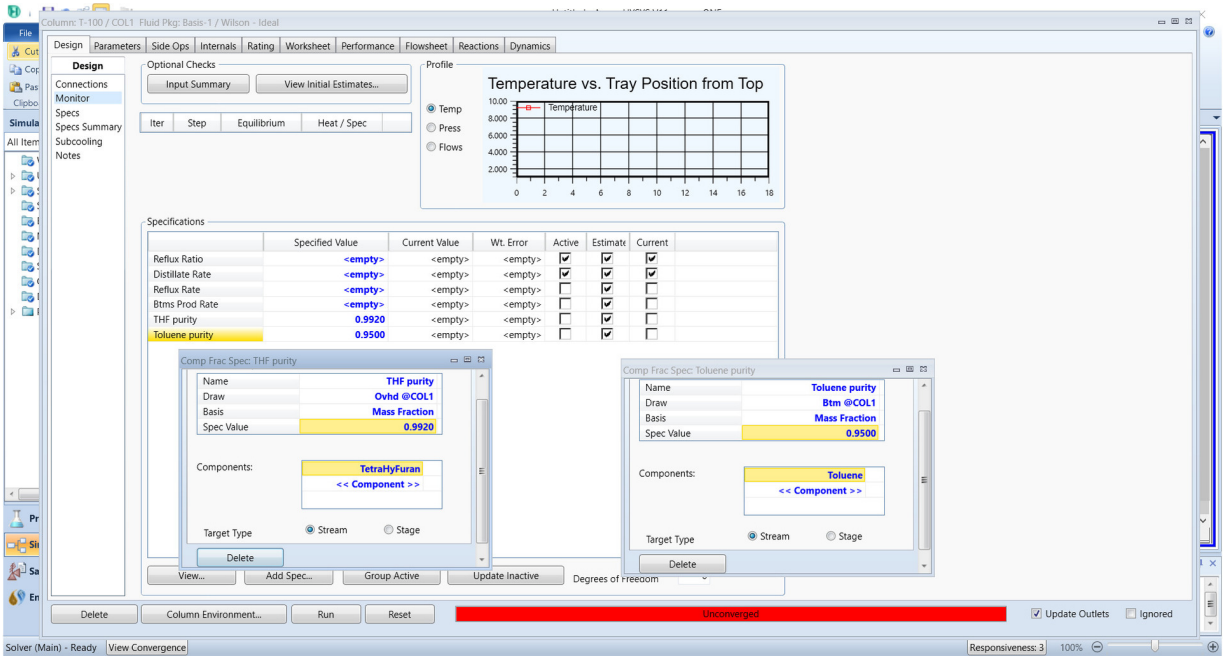
Click “Add Spec.”.
Click “Column Component Fraction” and “Add Spec.”.

Figure 12.14: Distillation column



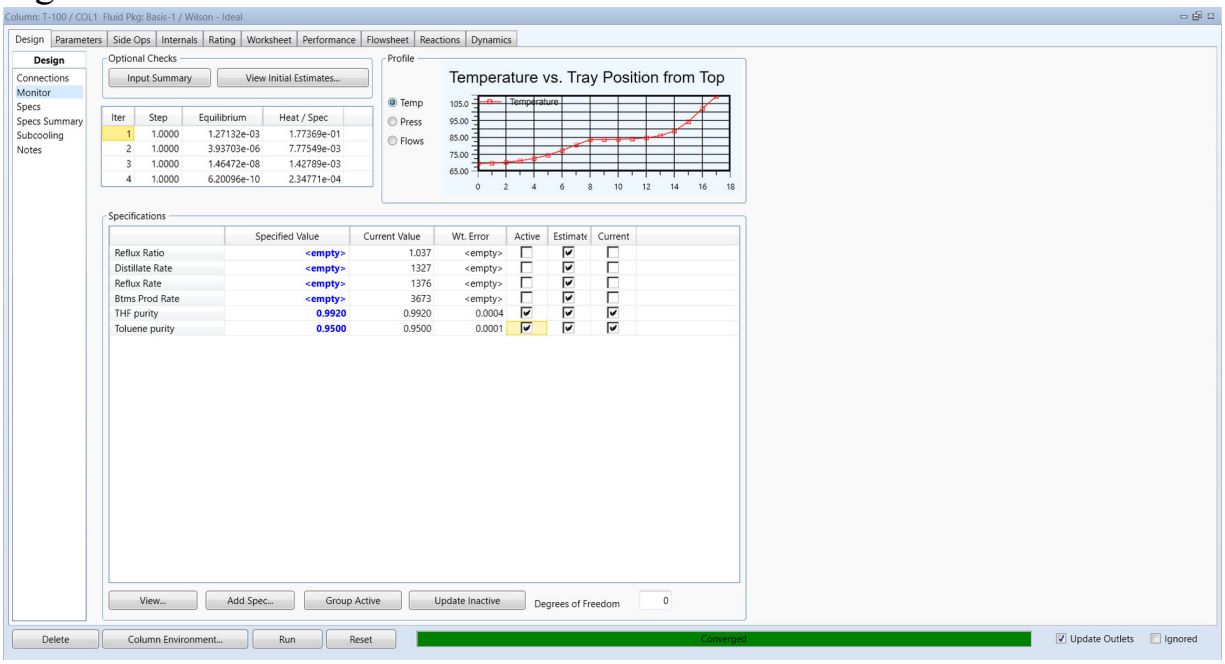
Provide specs. for THF and Toluene.

Figure 12.15: Distillation column



Select “THF Purity” and “Toluene Purity” as active specifications.

Figure 12.16: Distillation column

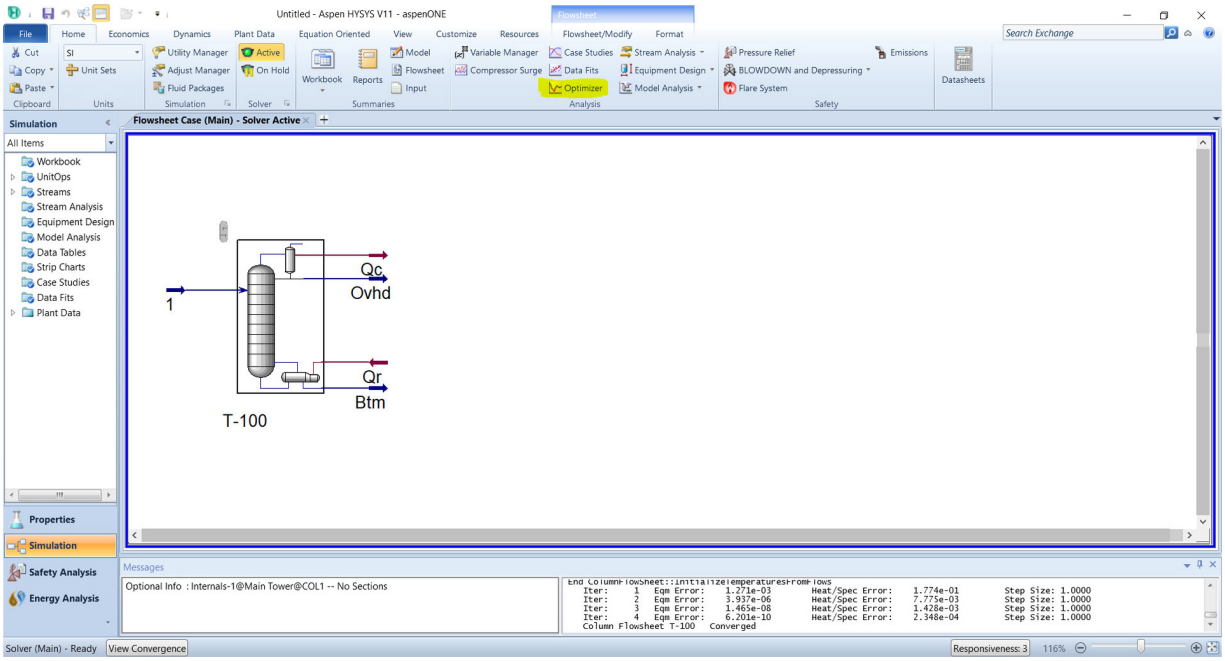


Optimization

We need to check if the operating conditions are optimum or not. The Variables to check are: THF Purity & Toluene Purity.

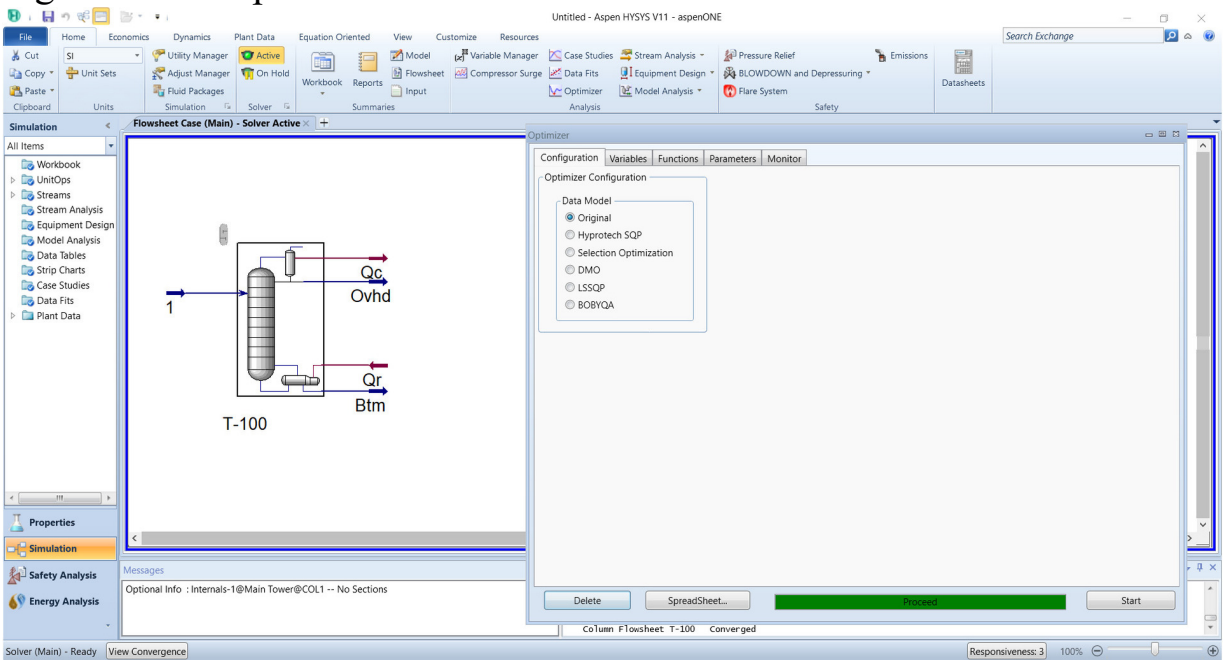
Using the Optimizer tool.
Click “Home” >> “Optimizer”.

Figure 12.17: Optimizer



Select “Original” as “Data Model” and click “Variable”.

Figure 12.18: Optimizer



Click “Add”.

Figure 12.19: Optimizer

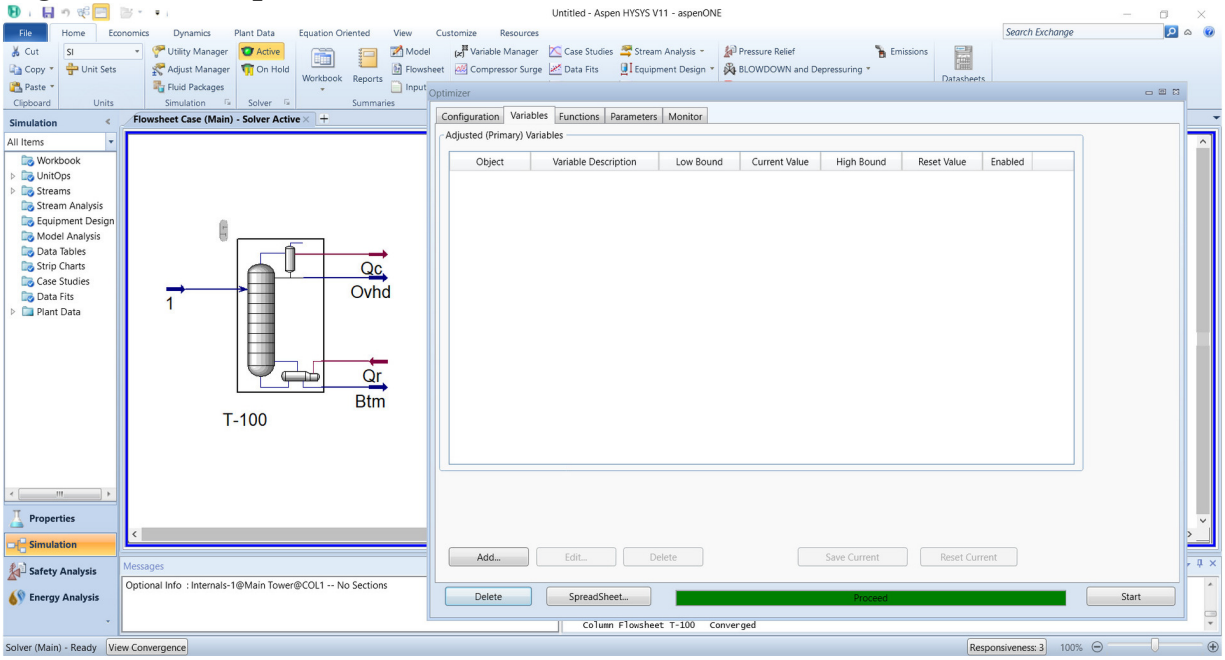
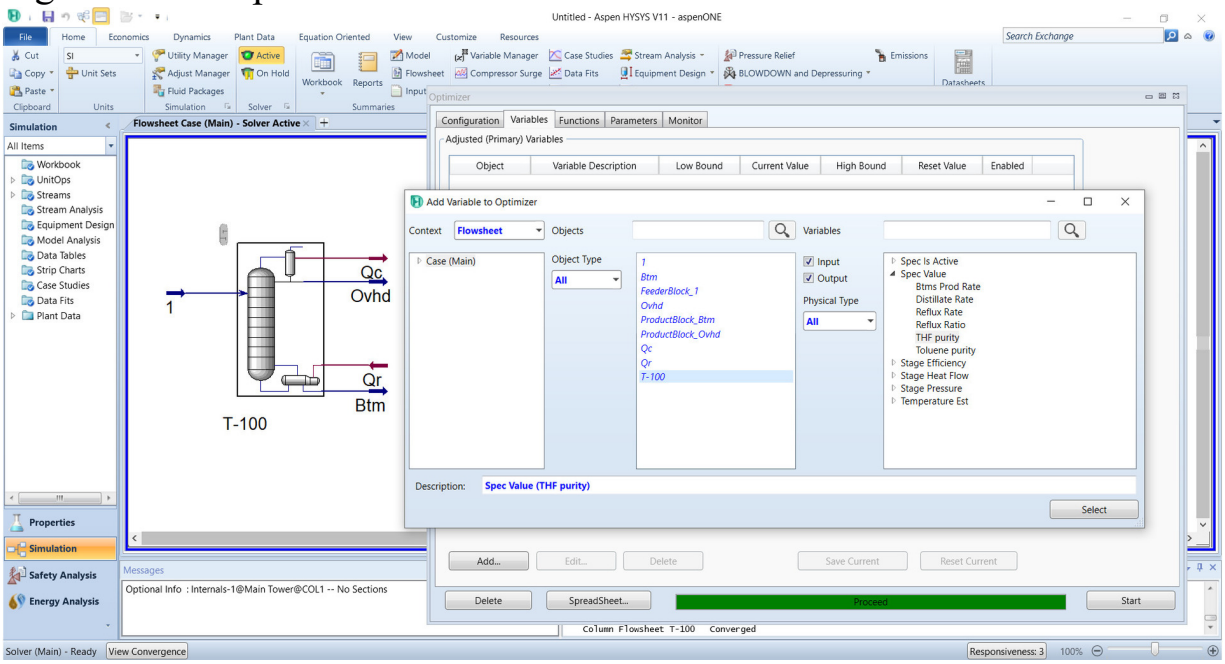
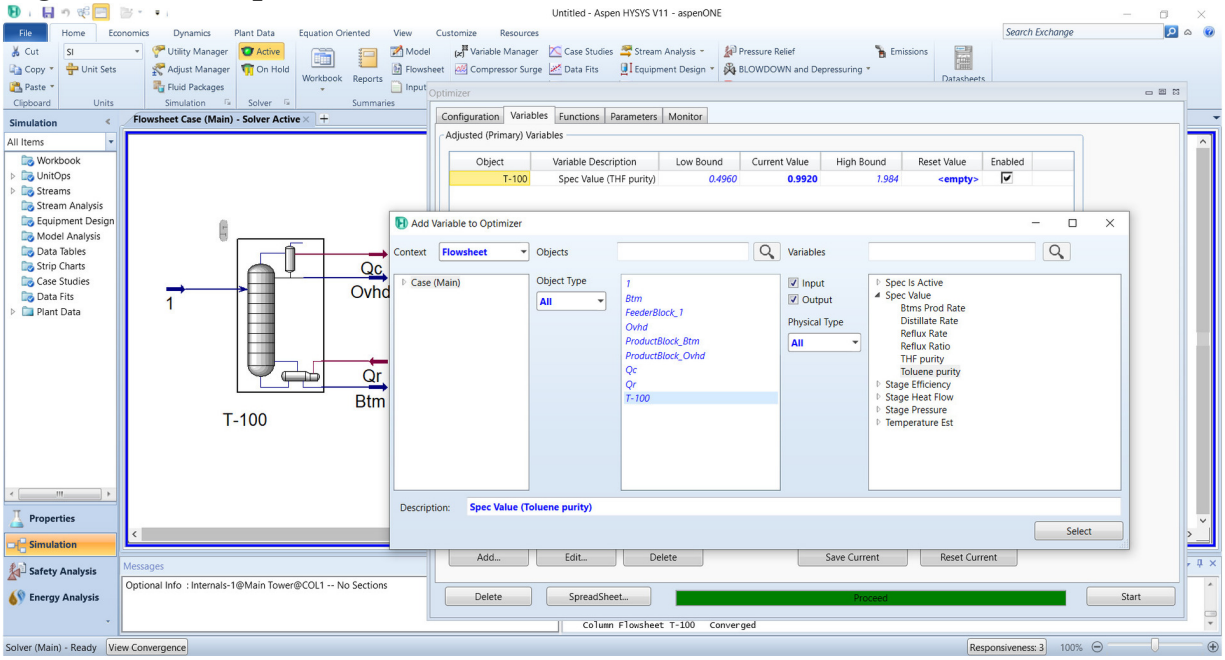


Figure 12.20: Optimizer



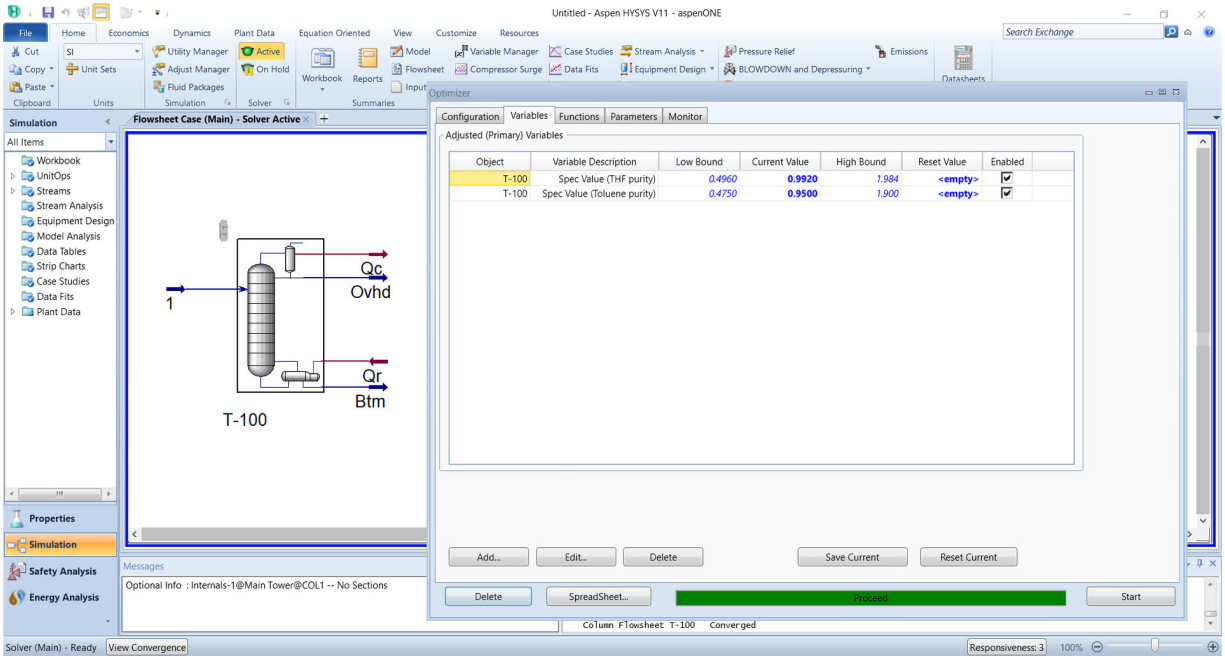
Click “Select”.
And do the same steps to add the other variable “Toluene Purity”.

Figure 12.21: Optimizer



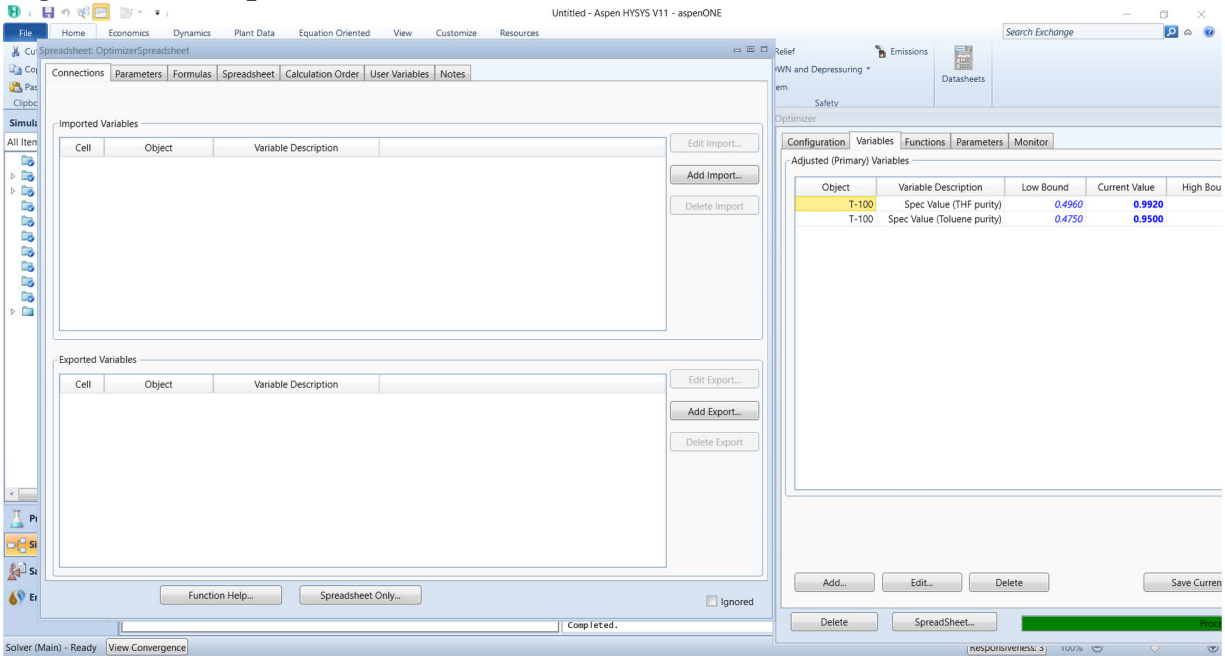
Click “Select”.
Use a range of 0.99 to 0.999 for the THF limit & 0.9 to 0.99 for the toluene.

Figure 12.22: Optimizer



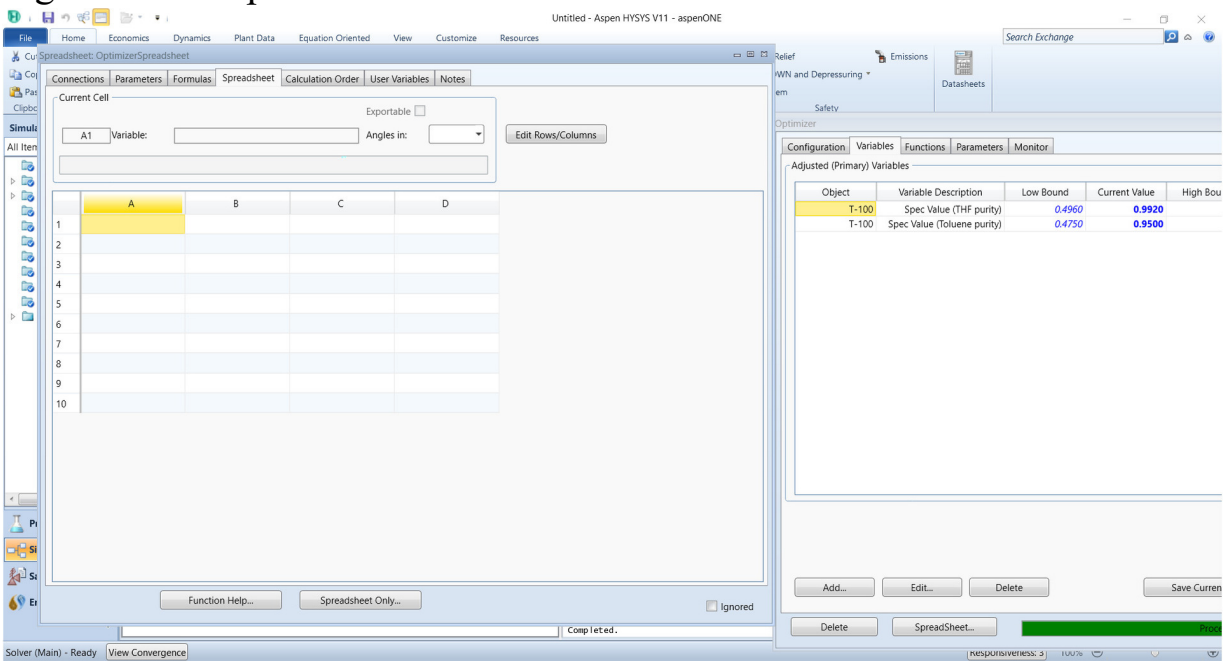
Click “Spreadsheet”.

Figure 12.23: Spreadsheet



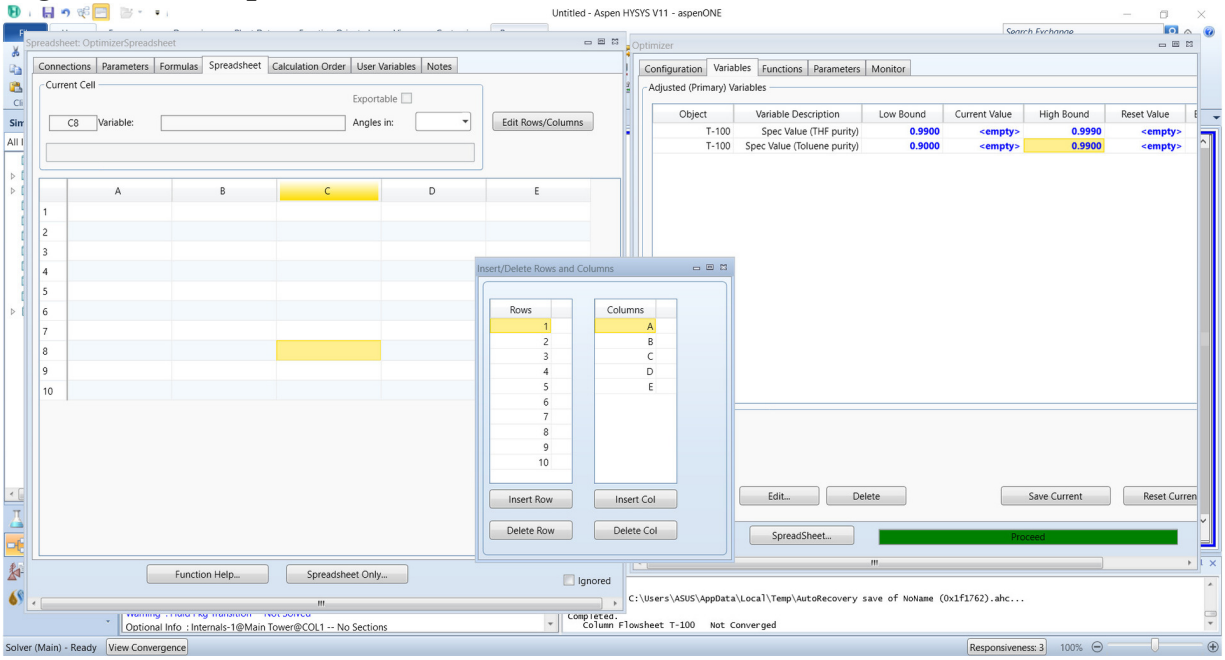
Select the “Spreadsheet” tab.

Figure 12.24: Spreadsheet



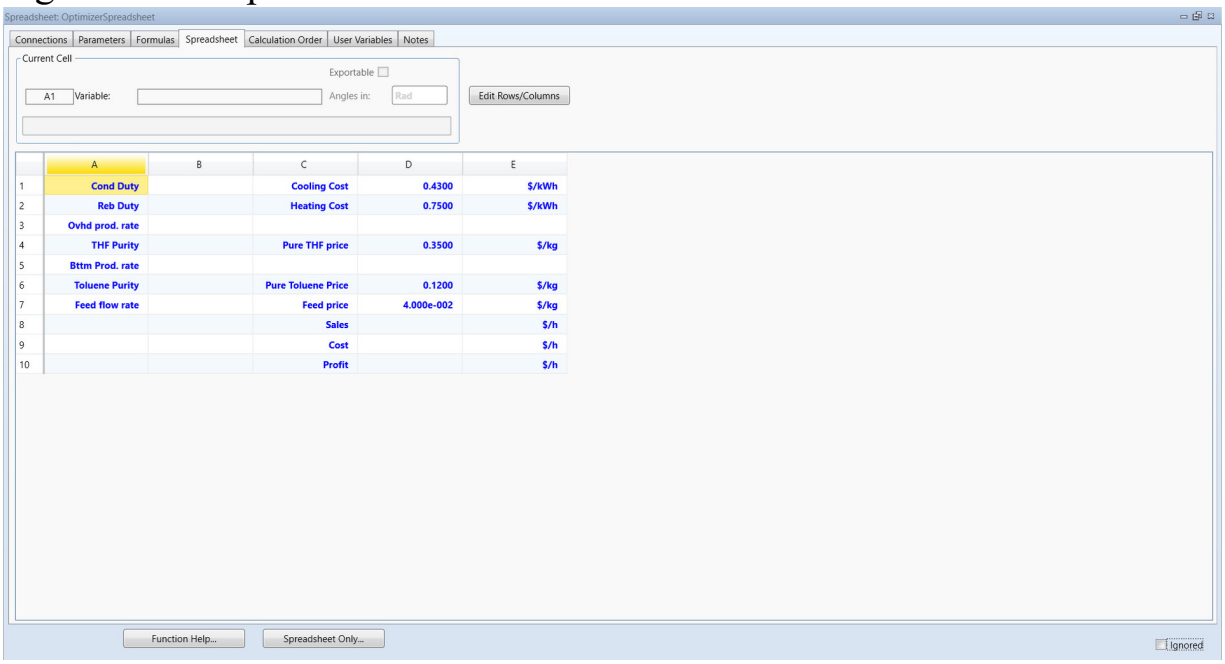
Insert a column.

Figure 12.25: Spreadsheet



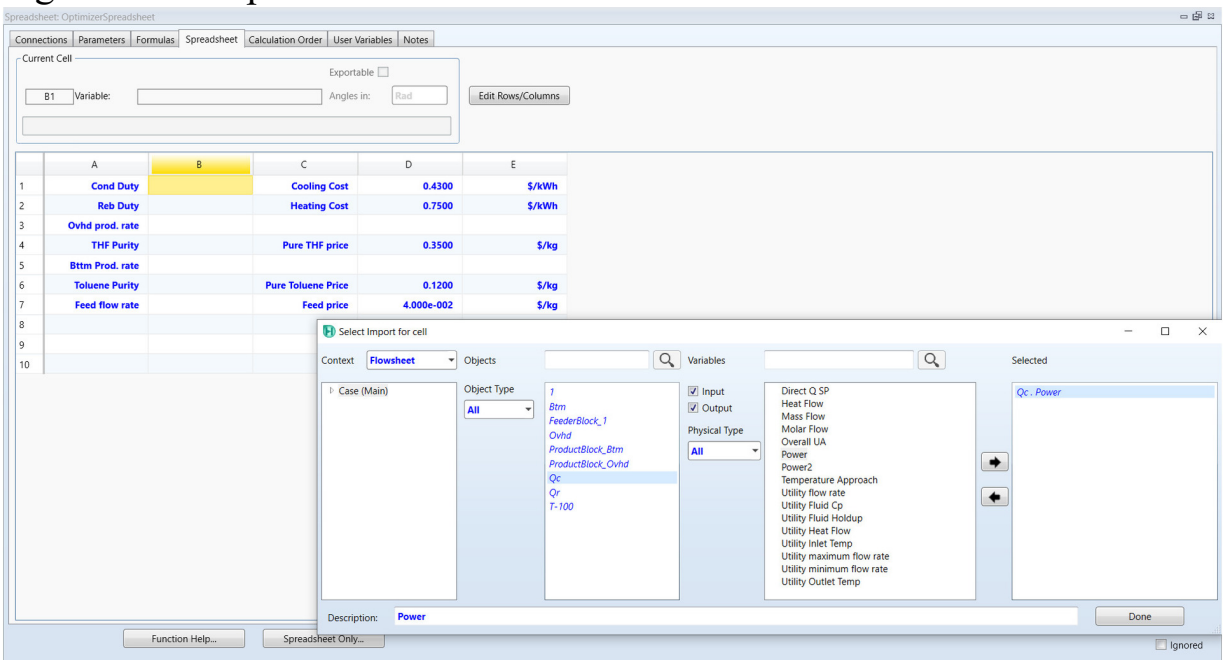
Type the text and data as given in below snapshot in columns A, C, D, and E.

Figure 12.26: Spreadsheet



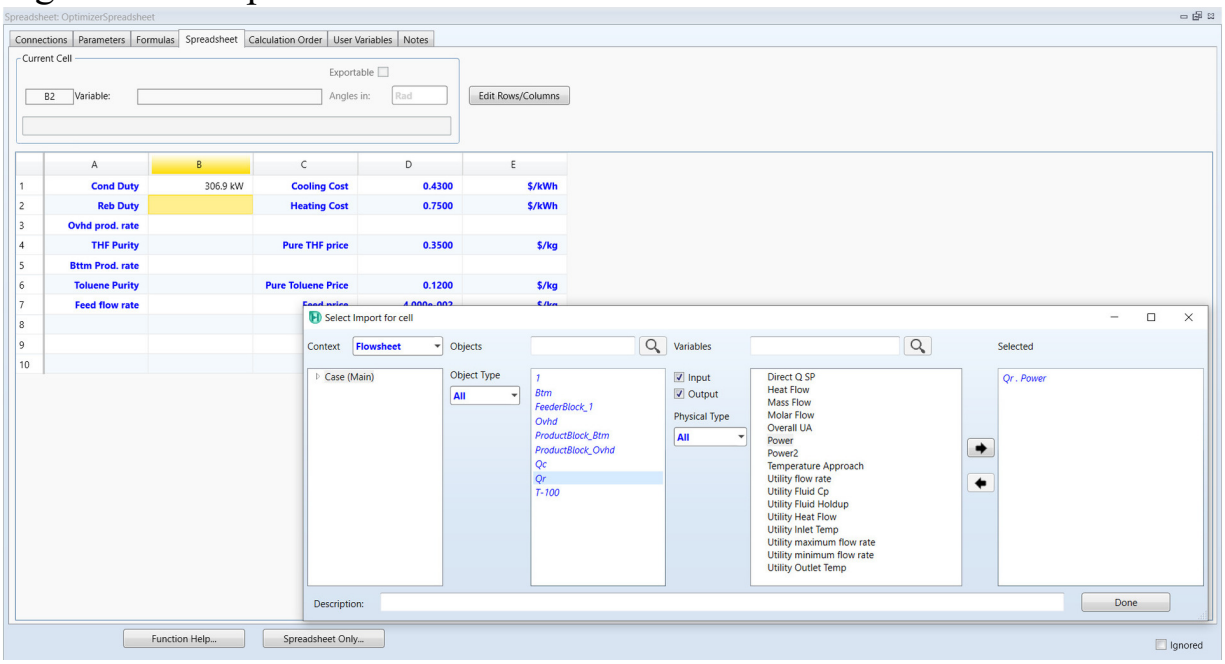
Right click “B1” >> “Import Variable” >> “Qc” >> “Power”.

Figure 12.27: Spreadsheet



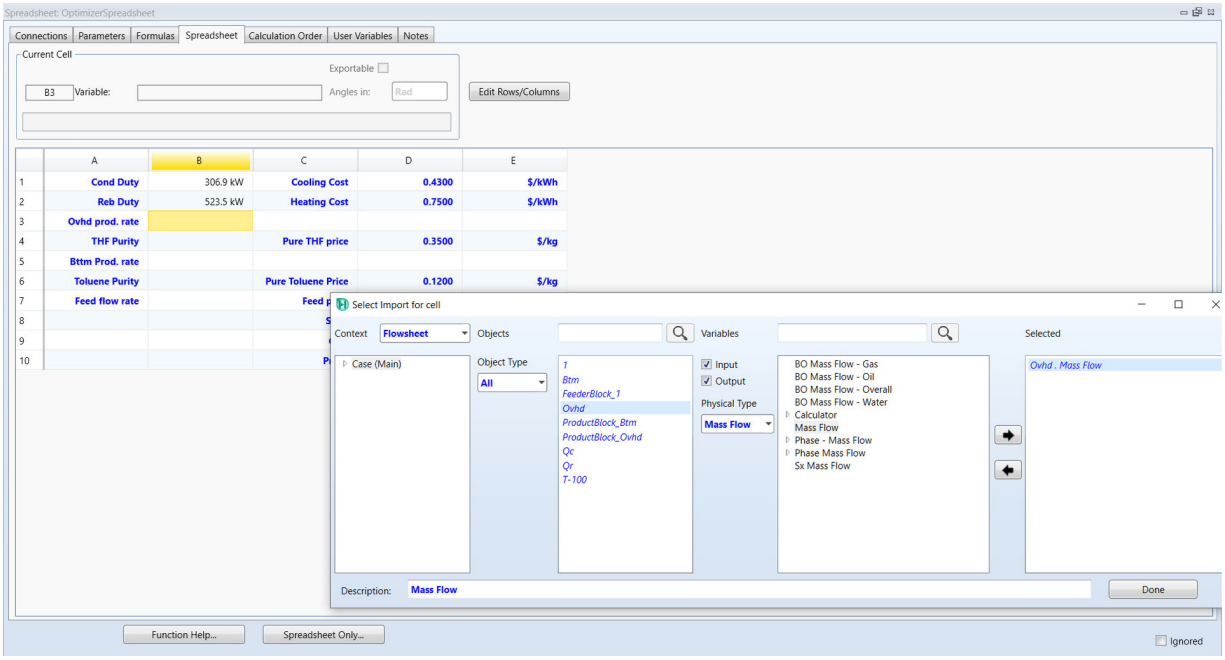
Click “Done”.
Right click “B2” >> “Import Variable” >> “Qr” >> “Power”.

Figure 12.28: Spreadsheet



Click “Done”.
Right click “B3” >> “Import Variable” >> “ovhd” >> “Mass Flow”.

Figure 12.29: Spreadsheet



The same procedure is to be used and other parameters are imported in the spreadsheet.

Figure 12.30: Spreadsheet

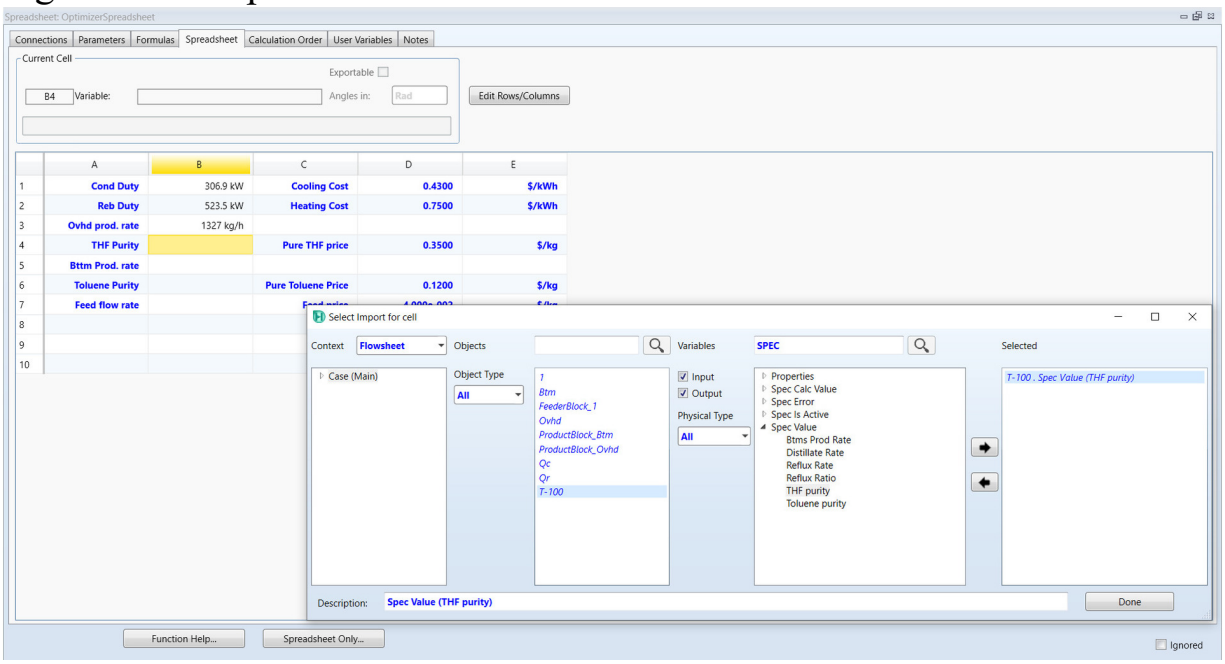


Figure 12.31: Spreadsheet

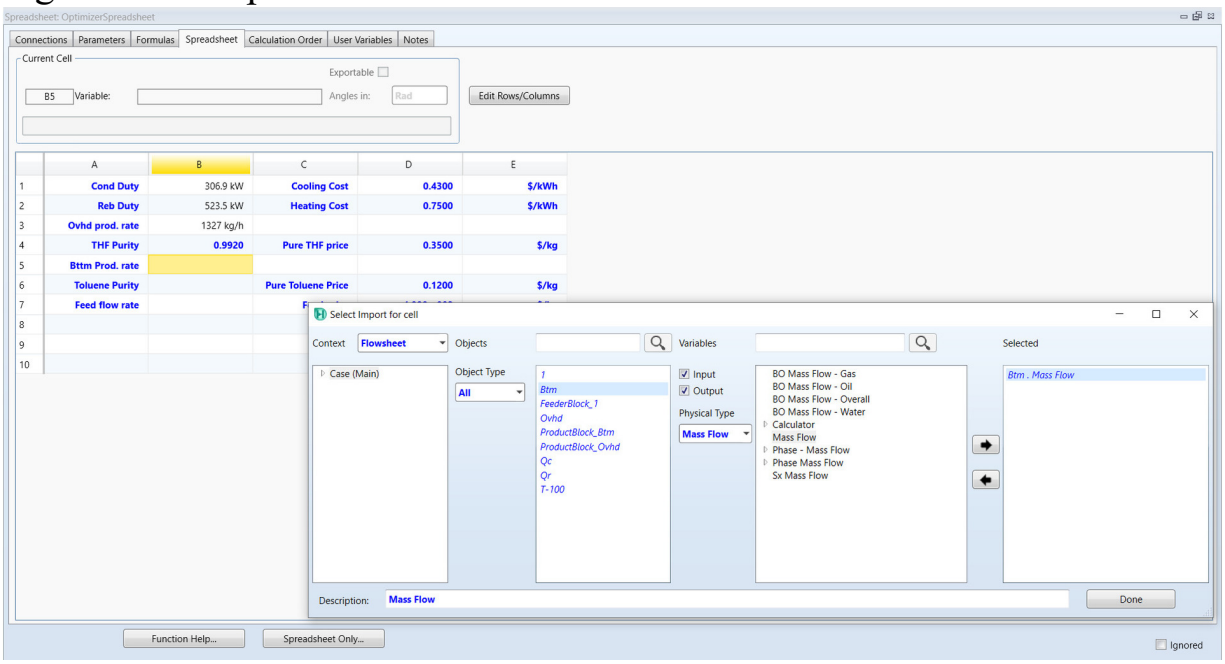


Figure 12.32: Spreadsheet

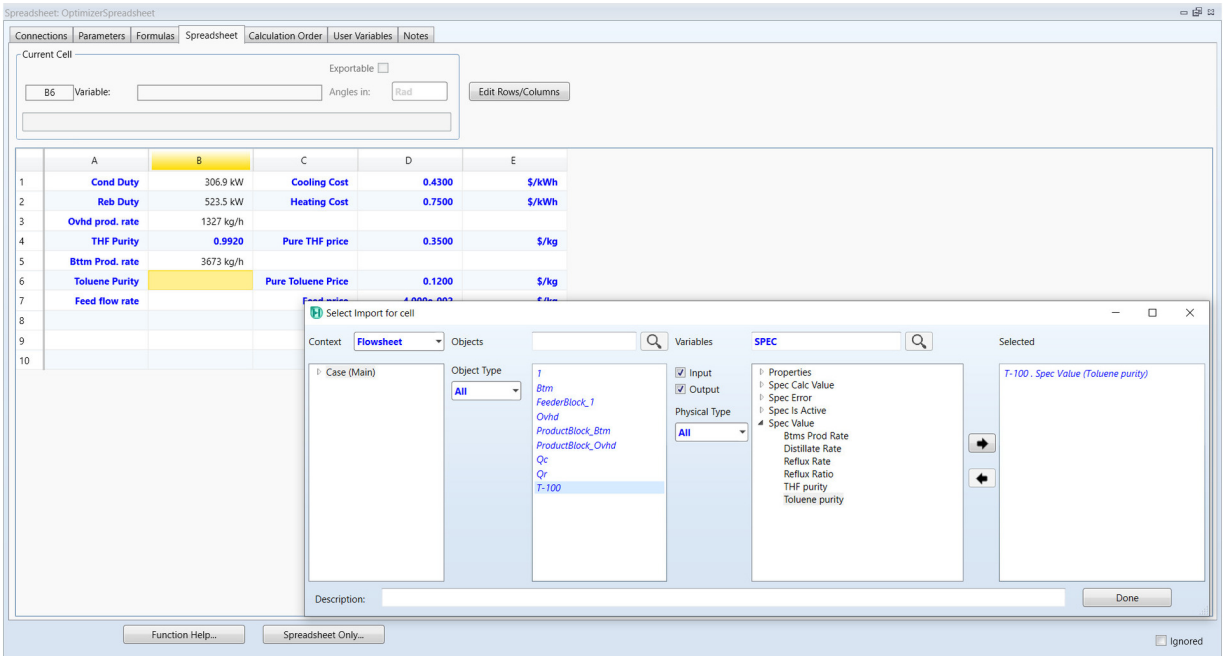


Figure 12.33: Spreadsheet

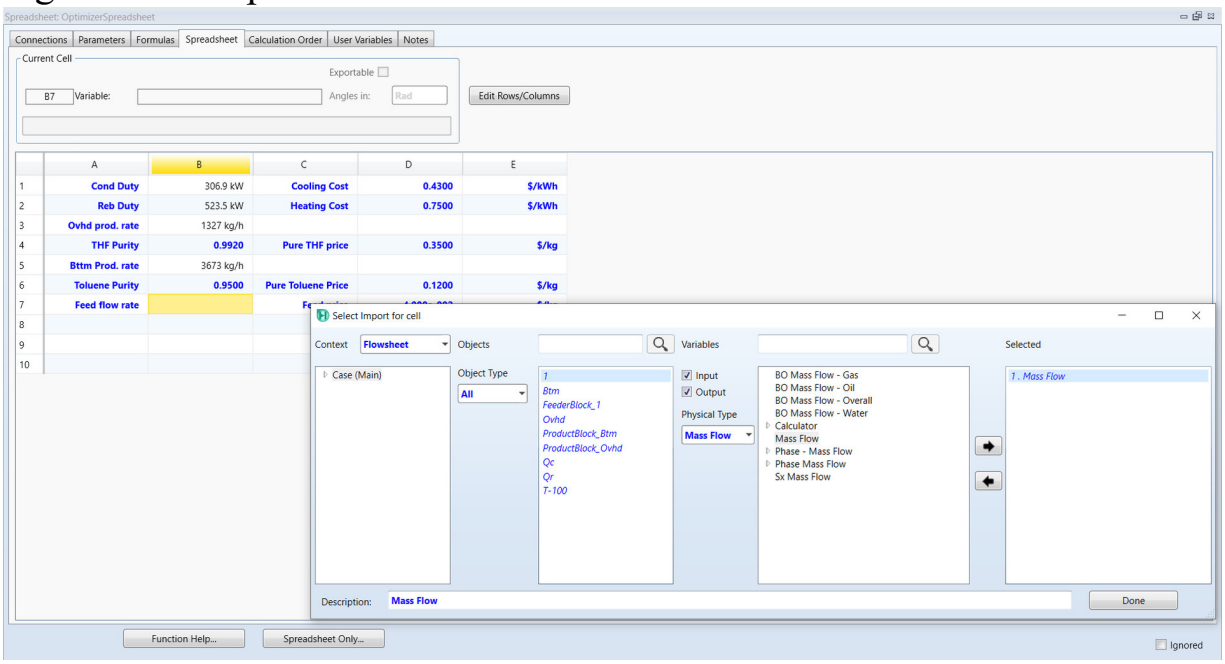
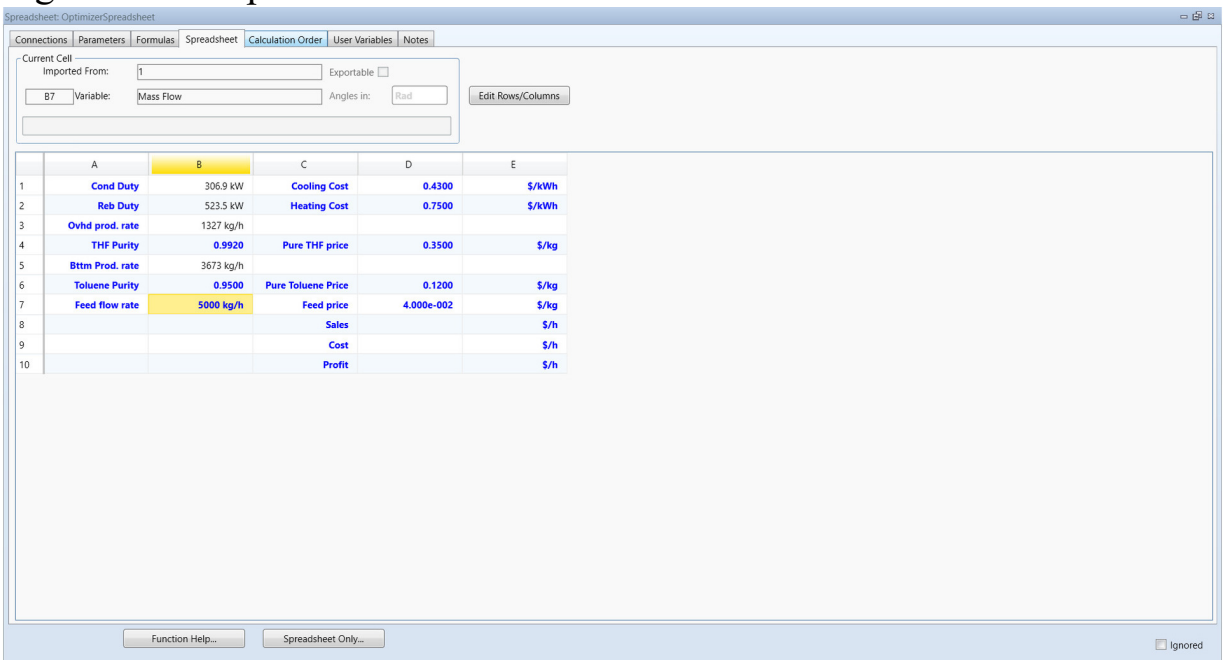


Figure 12.34: Spreadsheet



	A	B	C	D	E
1	Cond Duty	306.9 kW	Cooling Cost	0.4300	$/kWh
2	Reb Duty	523.5 kW	Heating Cost	0.7500	$/kWh
3	Ovhd prod. rate	1327 kg/h			
4	THF Purity	0.9920	Pure THF price	0.3500	$/kg
5	Bttm Prod. rate	3673 kg/h			
6	Toluene Purity	0.9500	Pure Toluene Price	0.1200	$/kg
7	Feed flow rate	5000 kg/h	Feed price	4.000e-002	$/kg
8			Sales		$/h
9			Cost		$/h
10			Profit		$/h

Provide Sales, Cost, and Profit formula in cell D8, D9, and D10 respectively.

Figure 12.35: Spreadsheet

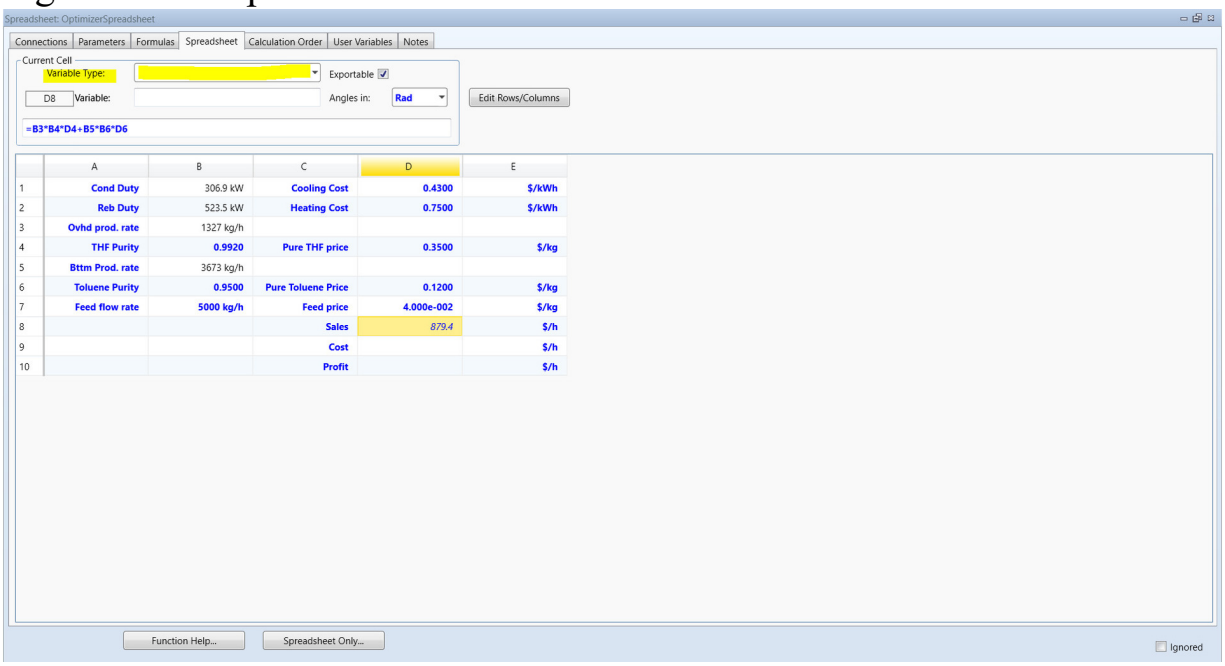
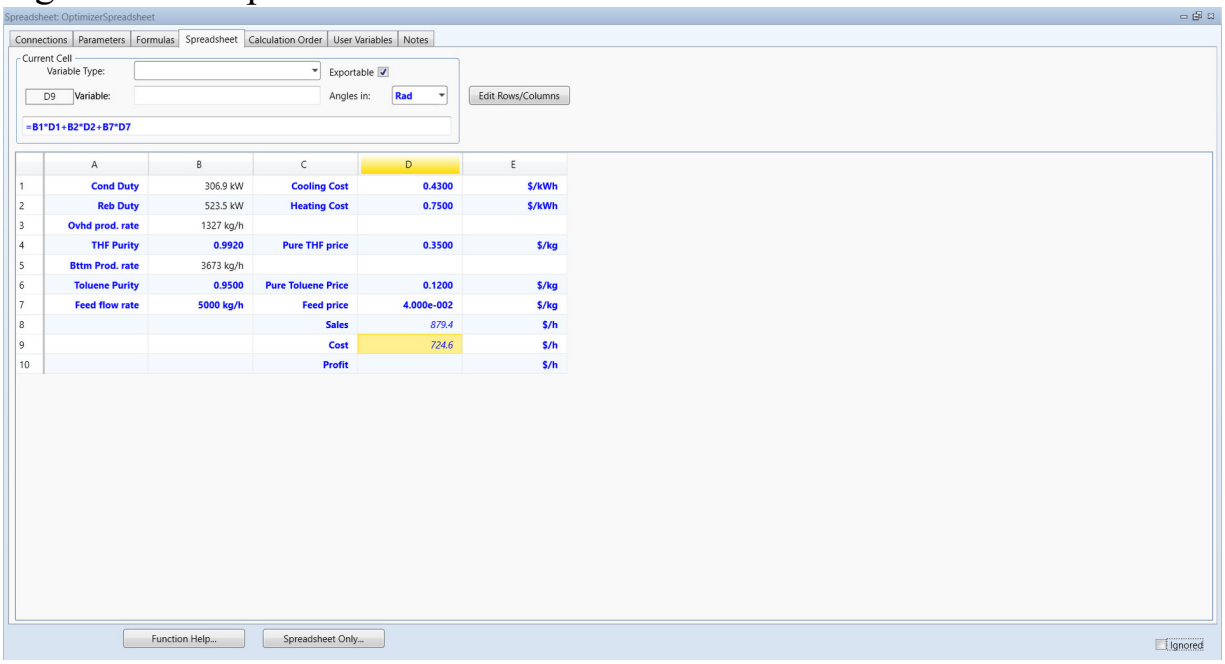
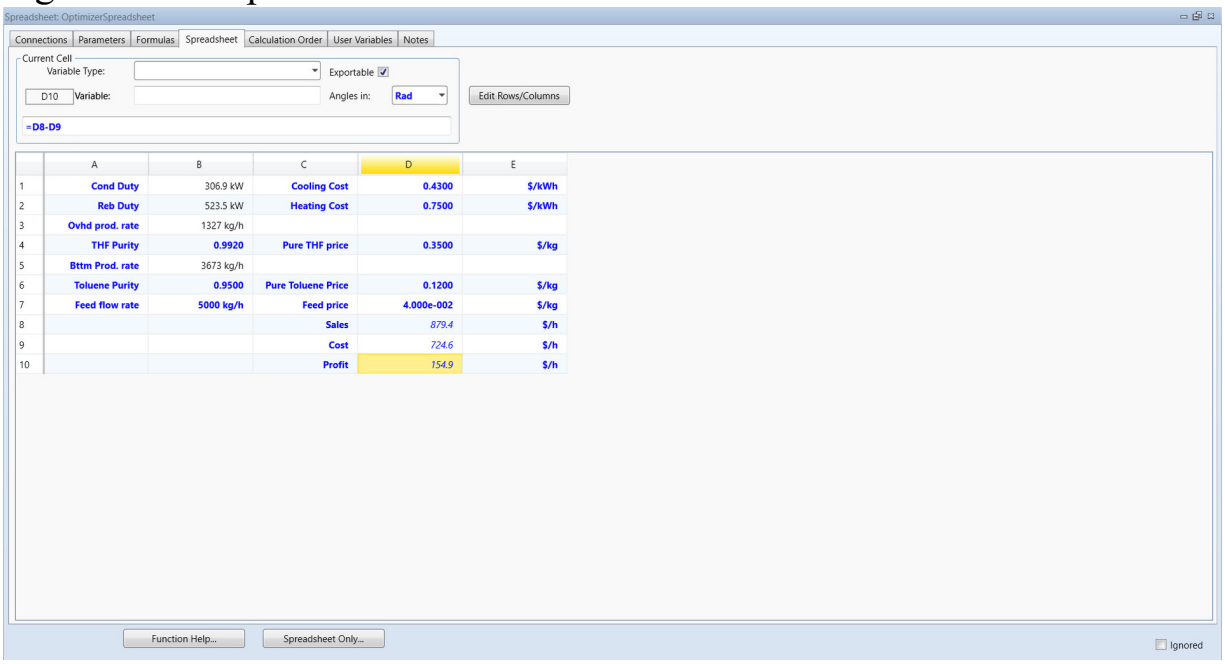


Figure 12.36: Spreadsheet



	A	B	C	D	E
1	Cond Duty	306.9 kW	Cooling Cost	0.4300	$/kWh
2	Reb Duty	523.5 kW	Heating Cost	0.7500	$/kWh
3	Ovhd prod. rate	1327 kg/h			
4	THF Purity	0.9920	Pure THF price	0.3500	$/kg
5	Bttm Prod. rate	3673 kg/h			
6	Toluene Purity	0.9500	Pure Toluene Price	0.1200	$/kg
7	Feed flow rate	5000 kg/h	Feed price	4.000e-002	$/kg
8			Sales	*879.4*	$/h
9			Cost	*724.6*	$/h
10			Profit		$/h

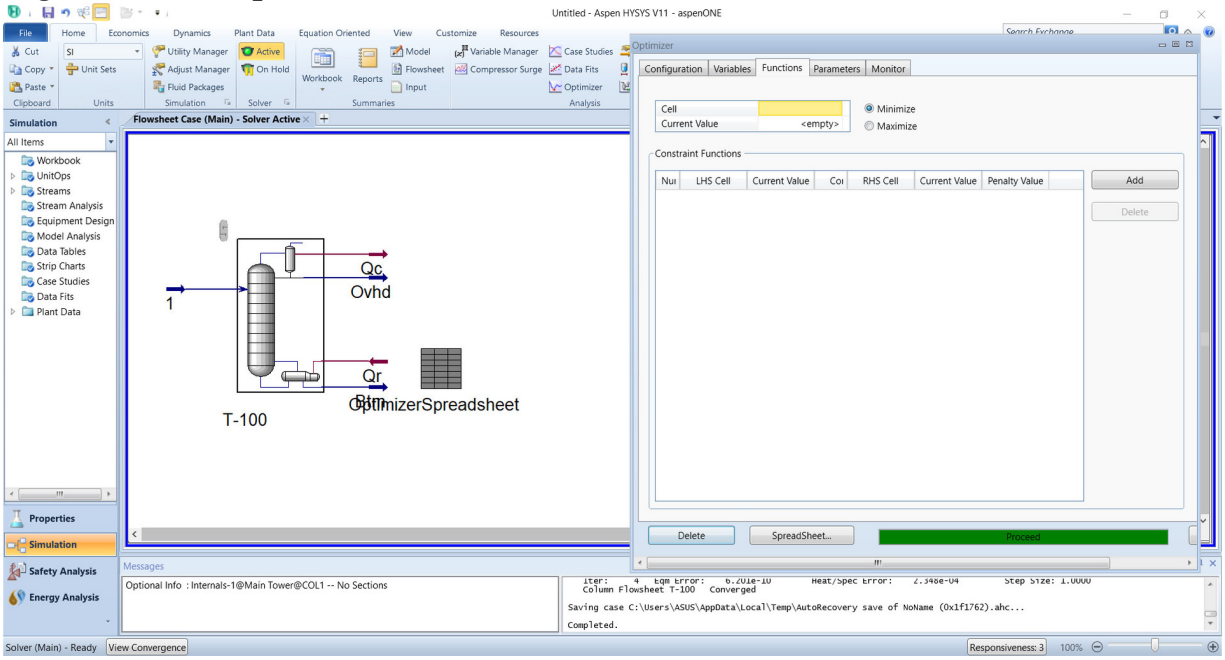
Figure 12.37: Spreadsheet



	A	B	C	D	E
1	Cond Duty	306.9 kW	Cooling Cost	0.4300	$/kWh
2	Reb Duty	523.5 kW	Heating Cost	0.7500	$/kWh
3	Ovhd prod. rate	1327 kg/h			
4	THF Purity	0.9920	Pure THF price	0.3500	$/kg
5	Bttm Prod. rate	3673 kg/h			
6	Toluene Purity	0.9500	Pure Toluene Price	0.1200	$/kg
7	Feed flow rate	5000 kg/h	Feed price	4.000e-002	$/kg
8			Sales	*879.4*	$/h
9			Cost	*724.6*	$/h
10			Profit	*154.9*	$/h

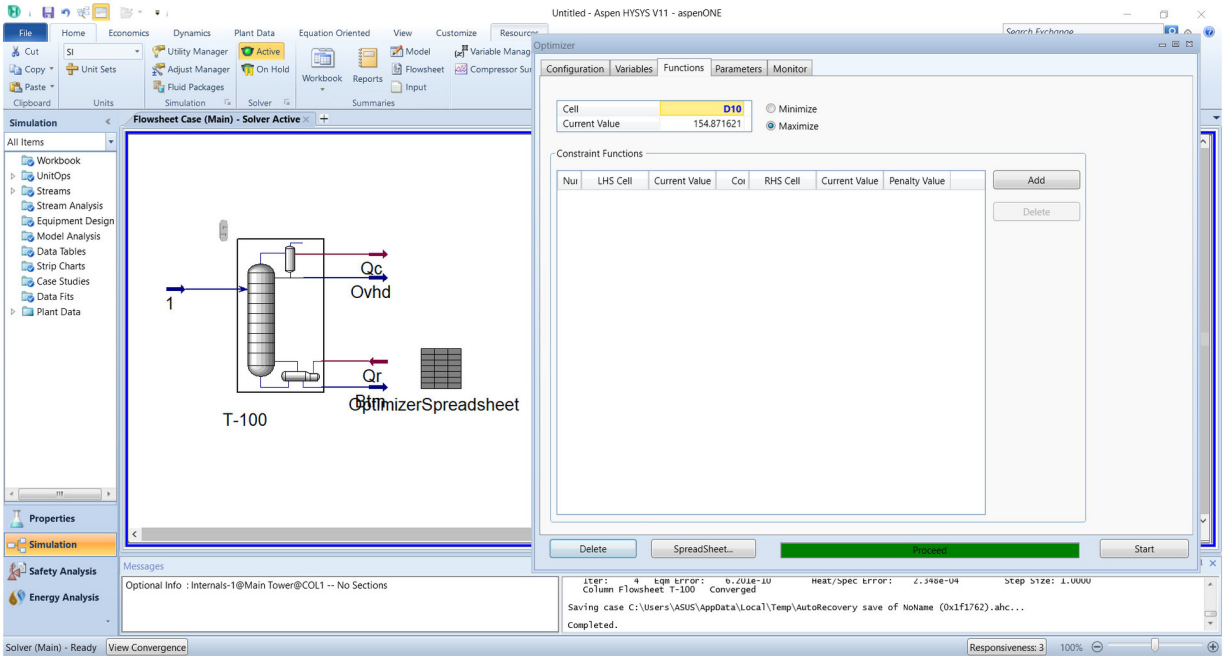
Close “Optimizer Spreadsheet”.
Click “Home” >> “Optimizer” >> Functions”.

Figure 12.38: Optimizer



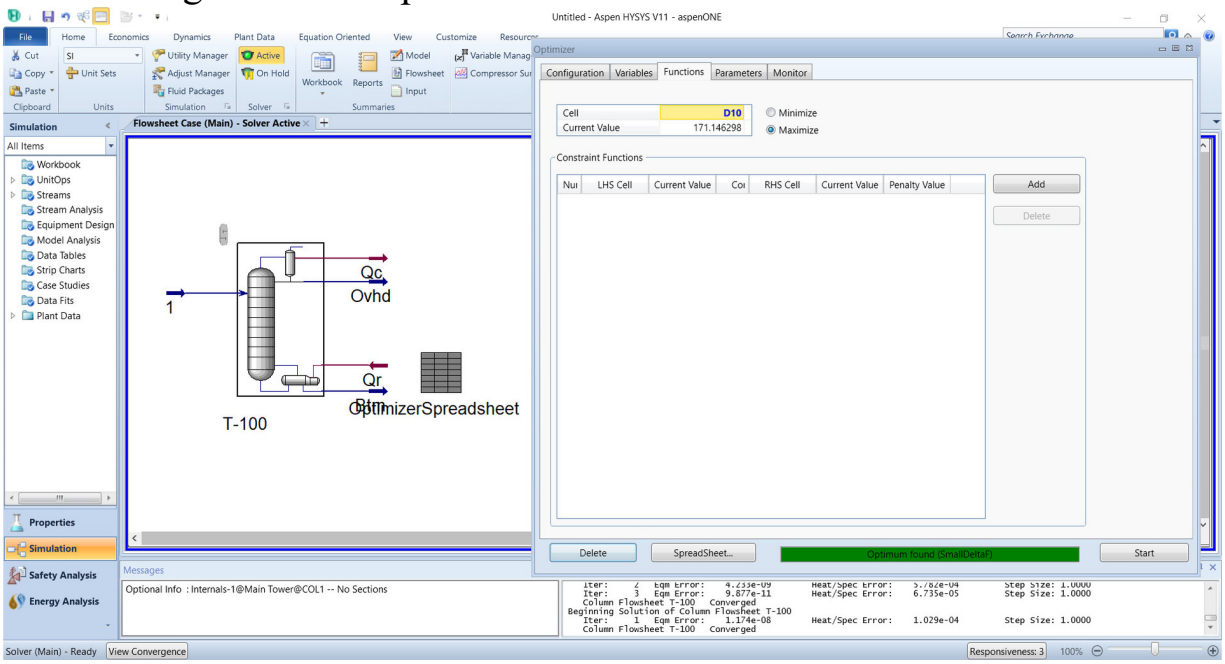
Select “D10” in “Cell”. Select “Maximize”.

Figure 12.39: Optimizer



Click “Start”.

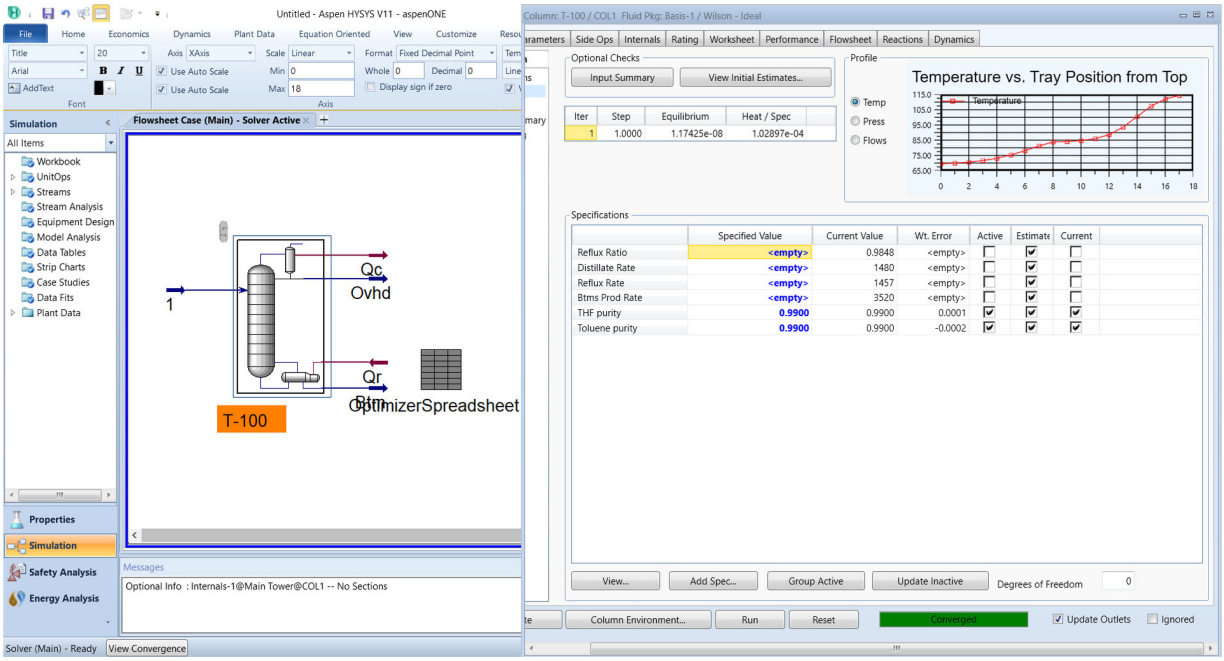
Figure 12.40: Optimizer



It can be noted that profit is increased from 155 $/h to 171 $/h using the optimization tool.

Now you can go to the monitor tab inside the column to see the optimum values for the THF & Toluene purities.

Figure 12.41: Optimizer



Exercises

a. Mr. Shihad is curious about what changes the optimizer tool has carried out to maximize the profit in the solved example. Can you figure it out?

b. Mr. Kartik is wondering what other problems can be resolved using the optimizer tool. Can you help him?

CHAPTER 13: PSV SIZING USING ASPEN HYSYS

Before attempting to size the PSV, we must know what is a PSV. Consider a vessel with a design pressure of 7.5 barg. Due to some reasons, if the vessel pressure increases beyond 7.5 barg, then it may lead to vessel leakage or it can burst also. To safeguard the vessel, a PSV is attached to the vessel top section. PSV is set at 7.5 barg. So, when the pressure reaches 7.5 barg, PSV opens and vessel pressure gets relieved. As pressure starts to go below 7.5 barg, PSV is closed.
Any equipment with design pressure equal to or more than 15 psig can be termed as a vessel. And the one with less than 15 psig is called a tank. A tank may have relief devices called as breather valve or PVRV and sizing are done as per API 2000.
The vessel may have a relief device called PSV or RD. There are the following 3 major standards that deal with PSV and RD.

- API 520 Part 1 : PSV and RD sizing equations are provided in this standard.
- API 521 : Depressurization, Flaring, liquid expansion is discussed in this standard.
- API 526 : The selection of PSV size that is available in the market is given in this standard.

Terminology

- **Operating pressure**

It is nothing but the normal operating pressure of the vessel.
In our example, let's assume operating pressure to be 5 barg. The maximum operating pressure (MOP) is usually taken at a 1.7 bar above normal operation. So MOP will be 6.7 barg. Vessels are often overdesigned relative to the maximum operating pressure, i.e. 10% over MOP. Hence design pressure value will be 7.37 barg or round figure 7.5 barg.

- **MAWP**

The maximum gauge pressure permissible at the top of a completed vessel in its normal operating position at the designated coincident temperature specified for that pressure. The pressure is the least of the values for the internal or external pressure as determined by the vessel design rules for each element of the vessel using actual nominal thickness, exclusive of additional metal thickness allowed for corrosion and loadings other than pressure. The maximum allowable working pressure is the basis for the pressure setting of the pressure relief devices that protect the vessel. The MAWP is normally greater than the design pressure but can be equal to the design pressure when the design rules are used only to calculate the minimum thickness for each element and calculations are not made to determine the value of the MAWP.
MAWP could be 7.5 barg or even more depending upon actual shell thickness of the vessel.

- **Design pressure**

Pressure, together with the design temperature, is used to determine the minimum permissible thickness or physical characteristic of each vessel component as determined by the vessel design rules. The design pressure is selected by the user to provide a suitable margin above the most severe pressure expected during normal operation at a coincident temperature. This pressure may be used in place of the maximum allowable working pressure (MAWP) in all cases where the MAWP has not been established. The design pressure is equal to or less than the MAWP.
Design pressure here is calculated to be 7.5 barg.

- **Set pressure**

The inlet gauge pressure at which the pressure relief device is set to open under service conditions.
Set pressure in this example will be a 7.5 barg. As per API 520 Part 1, we could have set it at less than 7.5 barg (say at 7 barg) also, but we didn't, to avoid unnecessary popping of PSV and loss of material. But if the vessel is derated (due to aging leading to reduction of shell thickness), then one may choose to set the PSV at even lower pressure.
If there are 2 PSVs (both working) protecting a single vessel, then the first PSV can be set at 7.5 barg (max. 100% of design pressure) and second will be set at 7.875 barg (max. 105% of design pressure).

- **Accumulation**

The pressure increase over the maximum allowable working pressure of the vessel expressed in pressure units or as a percentage of maximum allowable working pressure (MAWP) or design pressure. Maximum allowable accumulations are established by applicable codes for emergency operating and fire contingencies.

Table 13.1: Accumulated pressure and set pressure guideline

Contingency	**Single Device Installations**		**Multiple Device Installations**	
	Maximum Set Pressure, %	**Maximum Accumulated Pressure, %**	**Maximum Set Pressure, %**	**Maxin Accumu Pressur**
Non-Fire Case				
First Relief Device	100	110	100	116
Additional Device(s)	-	-	105	116
Fire Case				
First Relief Device	100	121	100	121
Additional Device(s)	-	-	105	121

Supplemental Device	-	-	110	121
NOTE: All values are percentages of the maximum allowable working pres				

The above table from API 520 Part 1 can be used to determine allowable set pressure and maximum accumulated pressure. If there is a single PSV, then for non-fire (blocked outlet) case, the vessel shall be allowed to have a maximum pressure of 110% of design pressure. If design pressure is 7.5 barg, then maximum pressure allowable shall be 110% of 7.5 barg (8.25 barg).
For the fire case, maximum accumulated pressure shall be 121% of design pressure, irrespective of the number of PSVs.
In the above table, the only variable is design pressure. Hence it is advisable to prepare an excel sheet to calculate all the parameters given in the table w.r.to design pressure. So whenever PSV sizing has to be done, the same table can be referred.

- **Overpressure**

The pressure increase over the set pressure of the relieving device. Overpressure is expressed in pressure units or as a percentage of set pressure. Overpressure is the same as accumulation only when the relieving device is set to open at the maximum allowable working pressure of the vessel.

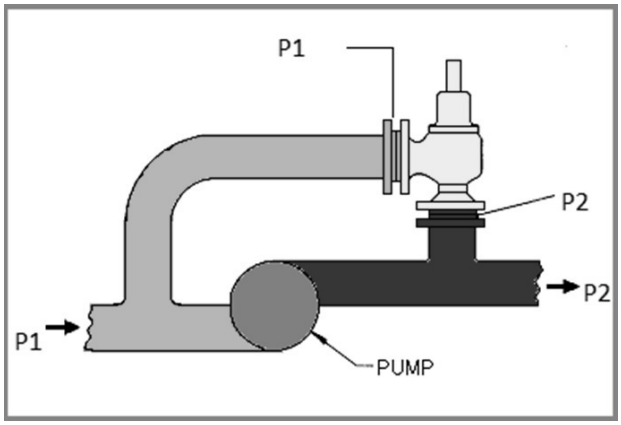
- **Blowdown**

The difference between the set pressure and the closing pressure of a pressure relief valve expressed as a percentage of the set pressure or in pressure units.
The closing pressure of PSV is a bit less than that of its set point. If the set pressure is 7.5 barg, then closing pressure may be 7.4 barg (depends on the PSV model). 0.1 bar shall be the blowdown.

- **Superimposed backpressure**

Pressure in discharge header before the valve opens.
Can be constant or variable.

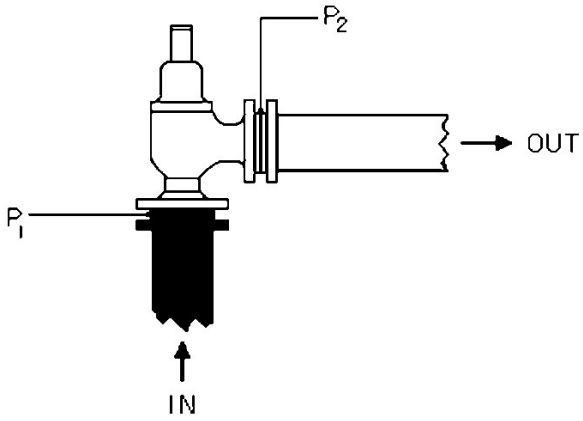
Figure 13.1: PSV at pump discharge



- Built-up back pressure
- Pressure in discharge header due to frictional losses after the valve opens

Total = Superimposed + Built-up

Figure 13.2: PSV



When to provide a PRV?

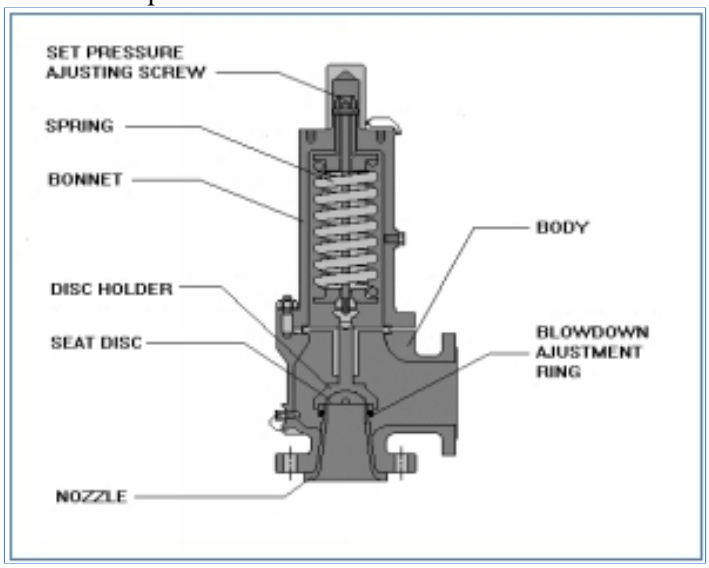
- All pressure vessels subject to overpressure shall be protected by a pressure-relieving device
- Liquid filled vessels or piping subject to thermal expansion must be protected by a thermal relief device
- Multiple vessels may be protected by a single relief device, provided there is a clear, unobstructed path to the device
- At least one pressure relief device must be set at or below the MAWP

Types of PSV
Conventional Spring-Loaded Safety Relief Valve

- Advantages
    - Most reliable type if properly sized and operated
    - Can be used in many services

- Disadvantages
    - Relieving pressure affected by back pressure
    - Susceptible to chatter if built-up back pressure is too high

Figure 13.3: PSV components

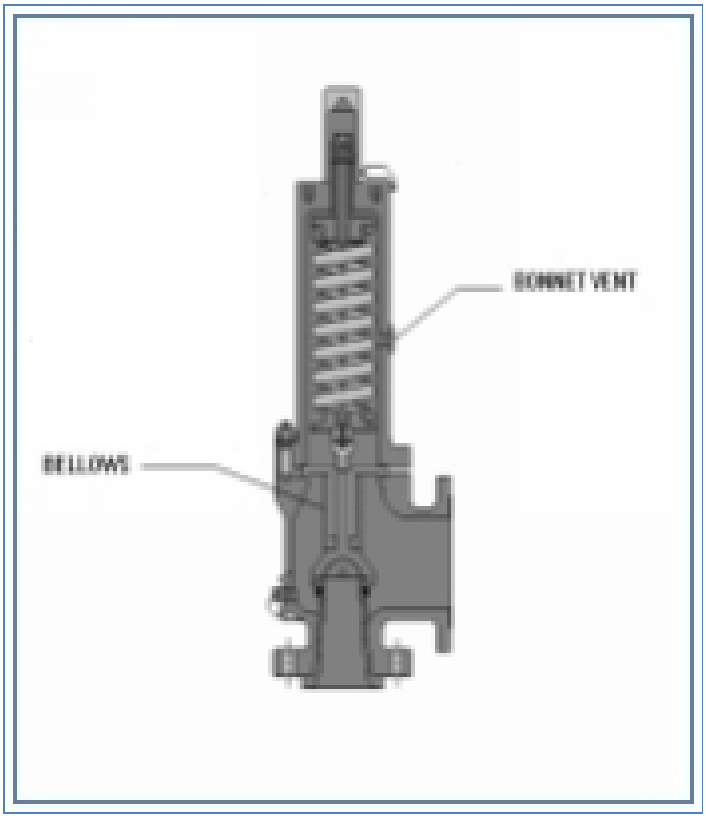


Balanced Bellows Spring Loaded Safety Relief Valve

- Advantages
    - Relieving pressure not affected by back pressure to a certain extent
    - Can handle higher built-up backpressure

- Disadvantages
    - Bellows susceptible to fatigue/rupture
    - May release flammables/toxics to atmosphere
    - Requires separate venting system

Figure 13.4: Balance bellow PSV



Piston Type Pilot Operated Safety Relief Valve

- Advantages
    - Relieving pressure not affected by back pressure

- Disadvantages
    - Pilot is susceptible to plugging
    - Potential for backflow

Figure 13.5: Pilot operated PSV

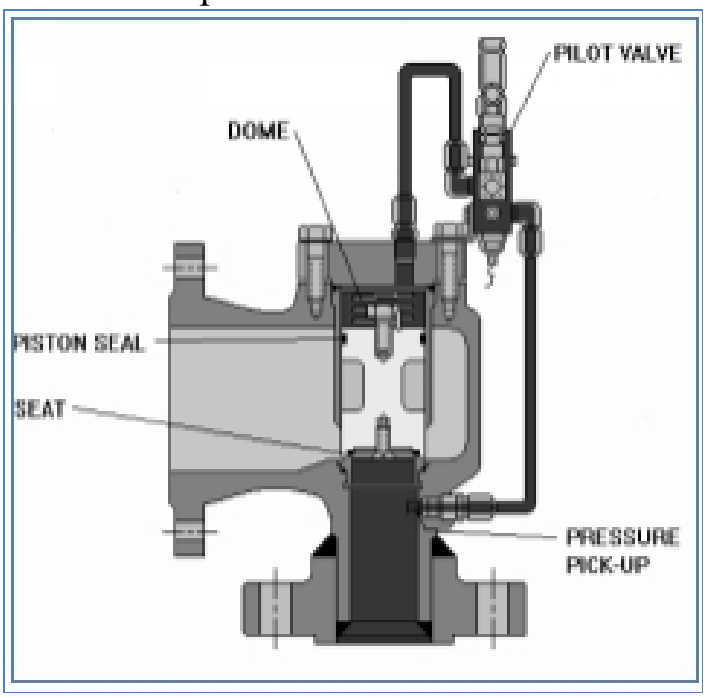


Table 13.2: PSV selection guideline

PSV type	When to use?
Conventional	When built-up backpressure is less than 10% of set pressure.
Balanced bellow	When total backpressure is between 10% to 50% of set pressure.
Pilot operated	When total backpressure is greater than 50% of set pressure.

Chatter

- Chattering is the rapid, alternating opening and closing of a PR Valve.
- Resulting vibration may cause misalignment, valve seat damage, and if prolonged, can cause mechanical failure of valve internals and associated piping.
- Chatter may occur in either liquid or vapor services.

Chatter causes:

- Excessive inlet pressure drop
- Excessive built-up backpressure
- Oversized valve
- Valve handling widely differing rates

Let's try to visualize the chattering problem caused by excessive pressure drop.

Figure 13.6: Chattering

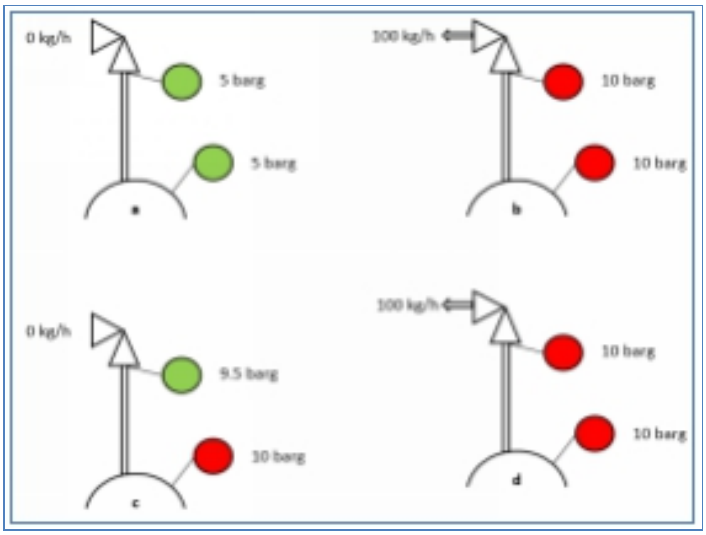
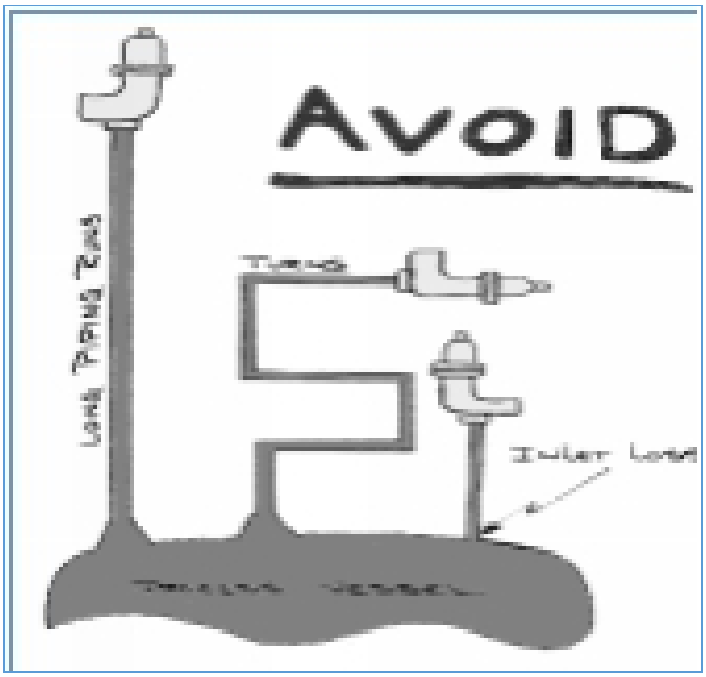


Figure a represents the normal operation, wherein the vessel operates at operating pressure. Hence the pressure at the pressure relief valve and the vessel is the same. Let's say that the operating pressure is 5 barg. As vessel pressure increases due to process upset, relief valve pressure will also increase. Relief valve will open at 10 barg, which the set pressure (figure b). So the pressure at the vessel and the relief valve is 10 barg. Let's assume that 100 kg/h vapor flows from the vessel to the relief valve. At the vessel, vapor pressure is 10 barg and due to pressure drop (say, 0.5 bar) in the relief valve inlet pipe, vapor will be left with 9.5 barg pressure. Since 9.5 barg is less than the relief valve set pressure, the relief valve will close and pressure at the relief valve will rise again (figure c). When pressure reaches sept pressure, then the relief valve opens again (figure d).

This cycle is repeated over and over. It causes vibration and damage to the relief valve, pipes, and equipment.

Figure 13.7: Chattering



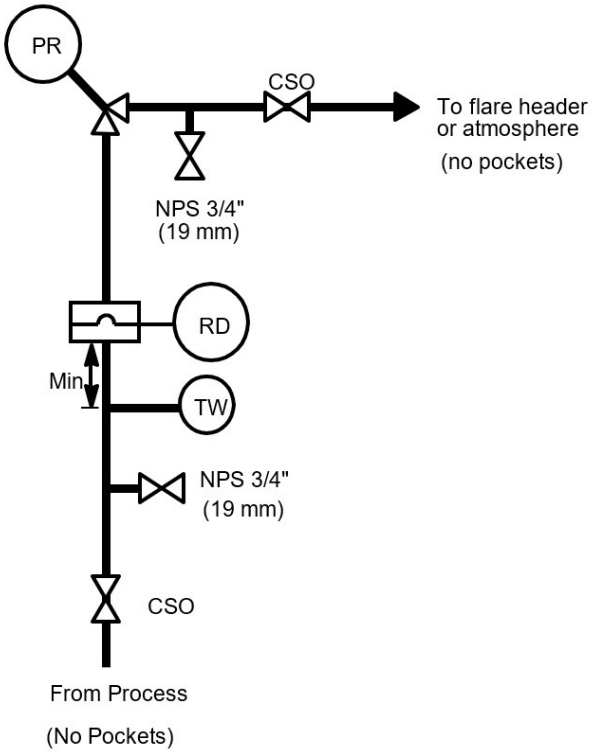
Chatter solution

- Avoid long piping runs
- Avoid bends/turns
- Minimize inlet loss
- Install smaller PRV
- Inlet line size must be at least equal to PRV inlet flange size
- Inlet piping should slope continuously upward from vessel to avoid traps
- Inlet piping should be heat traced if freezing liquids could occur
- A continual clean purge should be provided if coke/polymer formation or solids deposition

could occur

- Discharge line diameter must be at least equal to PRV outlet flange size
- Maximum discharge velocity should not exceed 70% of sonic velocity
- No check valves, orifice plates or other restrictions permitted
- Atmospheric discharge risers should have a drain hole
- The piping design must consider thermal expansion due to hot/cold release
- Auto-refrigeration and need for brittle fracture-resistant materials
- Closed discharge piping should slope continuously downward to header to avoid liquid traps

Figure 13.8: PSV and RD arrangement



Rupture disc

- A rupture disc is a thin diaphragm (generally a solid metal disc) designed to rupture (or burst) at a designated pressure.
- It is used as a weak element to protect vessels and piping against excessive pressure (positive or negative).
- A typical arrangement of RD + PSV is shown. Such arrangements are utilized when the process fluid is corrosive and it would be quite expensive to use PSV (alloy steel) without the aid of RD. But one can use alloy steel RD with carbon steel PSV. This makes the system quite cheap.

quiz

Q. **If a single safety valve is present only for fire, it is permissible to set it at 110% of the MAWP** . (T/F)

A. False. A single safety valve must be set no higher than the MAWP. Only if it is a second valve for a fire contingency may it be set at 105% of MAWP. And a third valve can be set at 110% of MAWP.

Q. **The highest allowable set pressure of any safety valve is the maximum allowable working pressure of the vessel being protected** . (T/F)

A. False. Under certain conditions, such as multiple valves, additional safety valves may be provided set at

pressures higher than the MAWP. However, at least one must be set no higher than MAWP.

Q. **The Design Pressure and the Maximum Allowable Working Pressure of a vessel are the same.** (T/F)
A. False. Design Pressure is a process design term that specifies the minimum pressure to which the vessel must be designed. The MAWP, on the other hand, is a mechanical design term. It goes with the vessel, i.e, it is the pressure on the vessel's nameplate and stays with the vessel no matter where the vessel is used. In practice, the two are often the same, but not necessarily.

Q. **An oversized safety valve can be vulnerable to the phenomenon known as chatter.** (T/F)
A. True.

Q. **Safety valve chatter in liquid service is potentially more serious than in vapor service. (** T/F)
A. True. Because of the liquid hammer effect.

Q. **For operating contingencies, the API Code allows the capacity of a single safety valve to be calculated at 110% of the MAWP.** (T/F)
A. True.

Q. **Under a fire contingency, the vessel is allowed to reach a higher pressure than under an operating contingency.** (T/F)
A. True. It is allowed to reach 121% of MAWP.

Q. **Accumulation means the same as blowdown.** (T/F)
A. False.

Q. **If there are two safety valves on a vessel, pressure during discharge is allowed to reach 116% of the MAWP.** (T/F)
A. True, assuming the second valve is set at 105% of MAWP as permitted by the code. With 10% accumulation,

maximum pressure becomes 110% of 105%, or (rounded) 116%.

Review of API 520 Part 1

This standard applies to the sizing and selection of pressure relief devices used in refineries and related industries for equipment that has a maximum allowable working pressure of 15 psig (103 kPag) or greater. The pressure relief devices covered in this standard are intended to protect unfired pressure vessels and related equipment against overpressure from operating and fire contingencies.

This standard includes basic definitions and information about the operational characteristics and applications of various pressure relief devices. It also includes sizing procedures and methods based on a steady-state flow of Newtonian fluids. One can refer sizing equations for PSV and RD sizing for gas (critical and subcritical), steam, liquid (with/without capacity certifications) from API 520 Part 1.

API 520 Part 2, however, deals with the installation procedure for the PSV and RD. It also provides the line sizing criteria for the PSV inlet line, which is that the pressure drop in the PSV inlet line should be less than 3% of set pressure.

Review of API 521

This standard applies to pressure-relieving and vapor depressuring systems. Although intended for use primarily in oil refineries, it is also applicable to petrochemical facilities, gas plants, liquefied natural gas (LNG) facilities, and oil and gas production facilities. The information provided is designed to aid in the selection of the system that is most appropriate for the risks and circumstances involved in various installations.

This standard specifies requirements and gives guidelines for the following:

- examining the principal causes of overpressure;
- determining individual relieving rates;
- selecting and designing disposal systems, including such parts as piping, vessels, flares, and vent stacks.

Pressure rise calculations due to hydraulic expansion and sizing calculations of TSV are provided in API 521.

Review of API 526

This standard is a purchase specification for flanged steel pressure-relief valves. Basic requirements are given for direct spring-loaded pressure-relief valves and pilot-operated pressure-relief valves as follows:

- orifice designation and area;
- valve size and pressure rating, inlet and outlet;
- materials;
- pressure-temperature limits;
- center-to-face dimensions, inlet, and outlet.

Table 13.3: Orifice designation

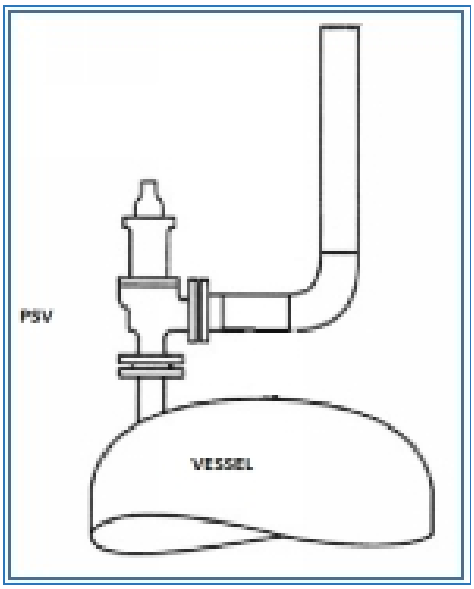
Designation	**Effective Orifice Area (in^2)**
D	0.110
E	0.196
F	0.307
G	0.503
H	0.785
J	1.287
K	1.838
L	2.853
M	3.60
N	4.34
P	6.38
Q	11.05
R	16.00
T	26.00

Simulation

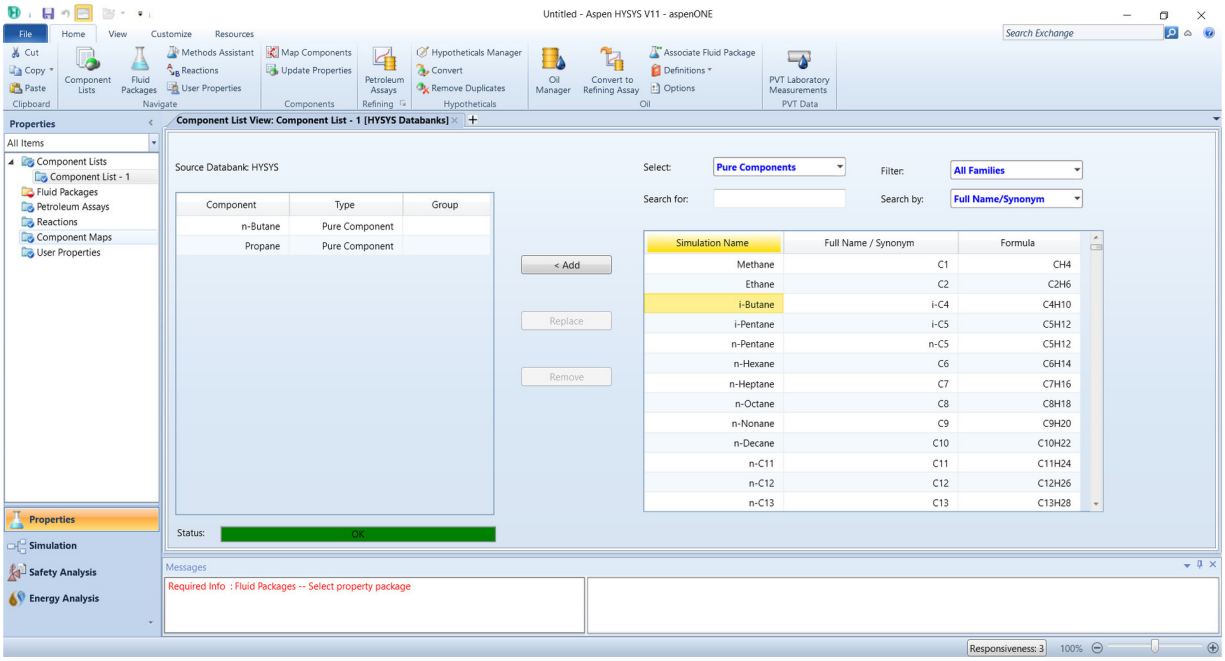
Let's solve an example. In this example, the following relief requirements are given:

- required hydrocarbon vapor flow, W, caused by an operational upset, of 50,000 lb/h;
- the hydrocarbon vapor is a 50/50 (by mole) mixture of n-butane and propane.
- the molecular weight of the vapor,
- relieving temperature, T, of 167 ^{o} F,
- PRV set at 75 psig, which is the design pressure of the equipment;
- backpressure of 0 psig
- overpressure of 10 %.

Figure 13.9: PSV



Find out PSV size using Aspen HYSYS.
Select components and fluid package. It may be noted that mostly for oil and gas problems, Peng Robinson fluid package is used. However, one can refer “Methods Assistant” tool to finalize the same as well.
Figure 13.10: Components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Navigate
Map Components
Update Properties
Components
Petroleum Assays
Refining
Hypotheticals Manager
Convert
Remove Duplicates
Hypotheticals
Oil Manager
Convert to Refining Assay
Associate Fluid Package
Definitions
Options
Oil
PVT Laboratory Measurements
PVT Data
Properties
All Items
Component Lists
Component List - 1
Fluid Packages
Petroleum Assays
Reactions
Component Maps
User Properties
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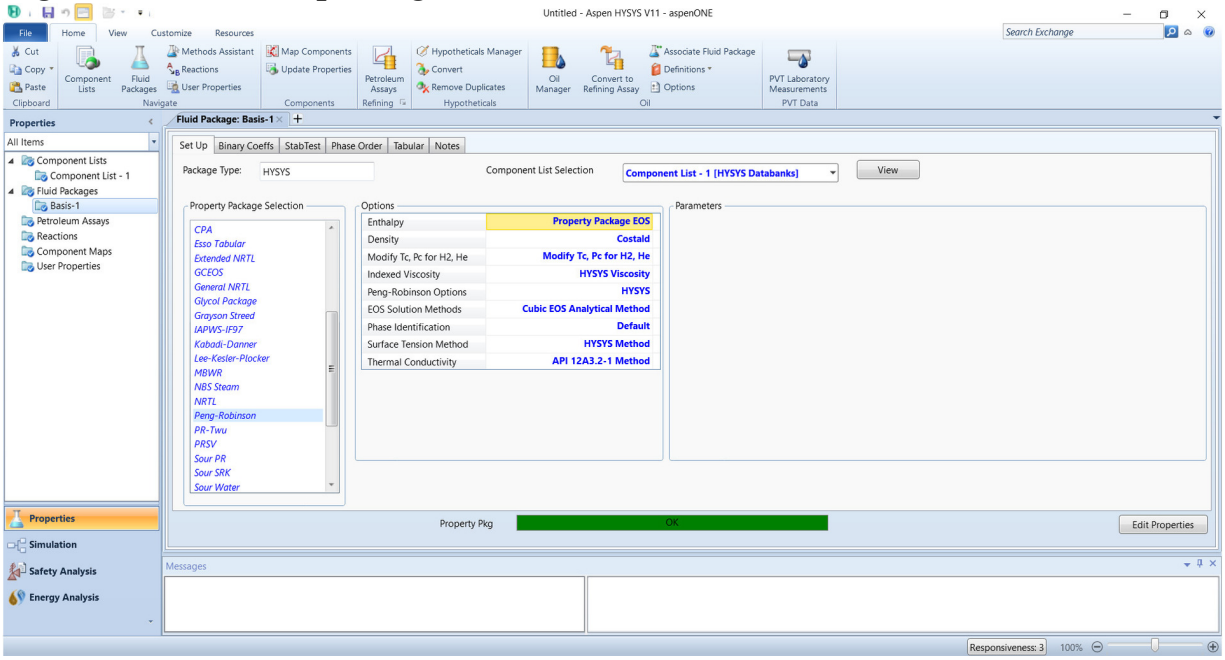
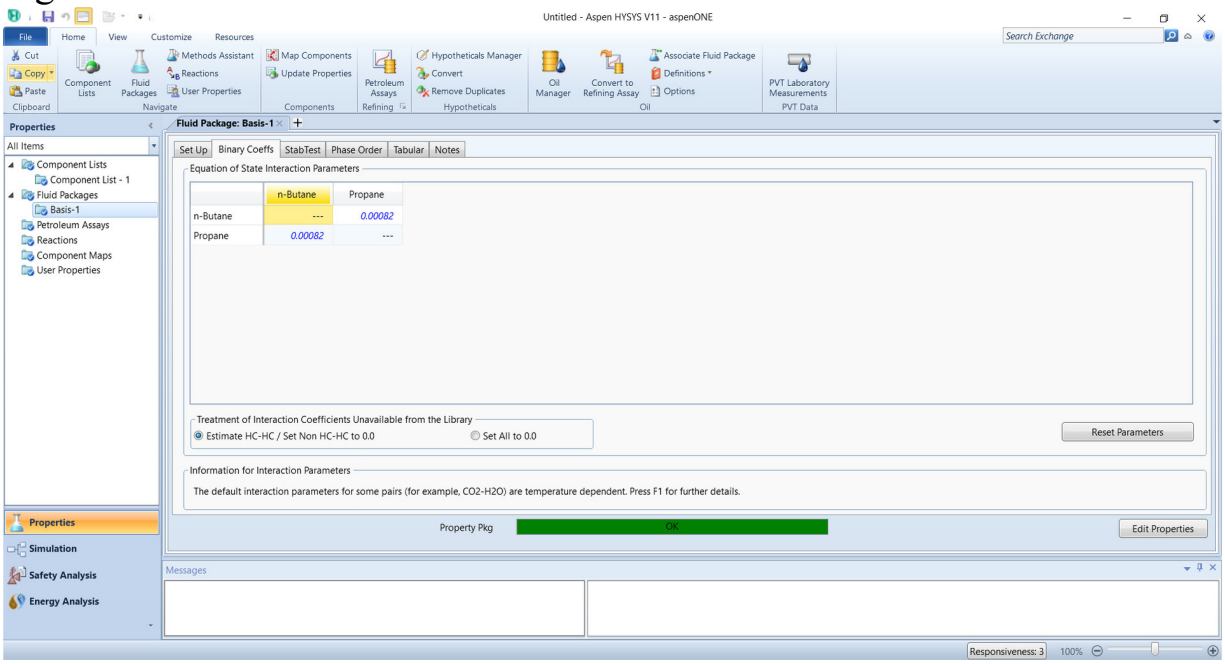
Source Databank: HYSYS
Component
Type
Group
n-Butane
Pure Component
Propane
Pure Component
< Add
Replace
Remove
Select:
Pure Components
Filter:
All Families
Search for:
Search by:
Full Name/Synonym
Simulation Name
Full Name / Synonym
Formula
Methane
C1
CH4
Ethane
C2
C2H6
i-Butane
i-C4
C4H10
i-Pentane
i-C5
C5H12
n-Pentane
n-C5
C5H12
n-Hexane
C6
C6H14
n-Heptane
C7
C7H16
n-Octane
C8
C8H18
n-Nonane
C9
C9H20
n-Decane
C10
C10H22
n-C11
C11
C11H24
n-C12
C12
C12H26
n-C13
C13
C13H28
Status:
OK
Properties
Simulation
Safety Analysis
Energy Analysis
Messages
Required Info : Fluid Packages -- Select property package
Responsiveness: 3
100%

Figure 13.11: Fluid package



It may also be checked if binary interaction parameters are available for the given components in the selected fluid package. In this case,

Figure 13.12: BIP



Enter Simulation environment and model a material stream. Provide pressure as 50 psig (normal operating pressure), temperature (167 ° F), 50,000 lb/h flow rate and composition 0.5-mole fraction each for n-Butane and Propane.

Figure 13.13: Material stream

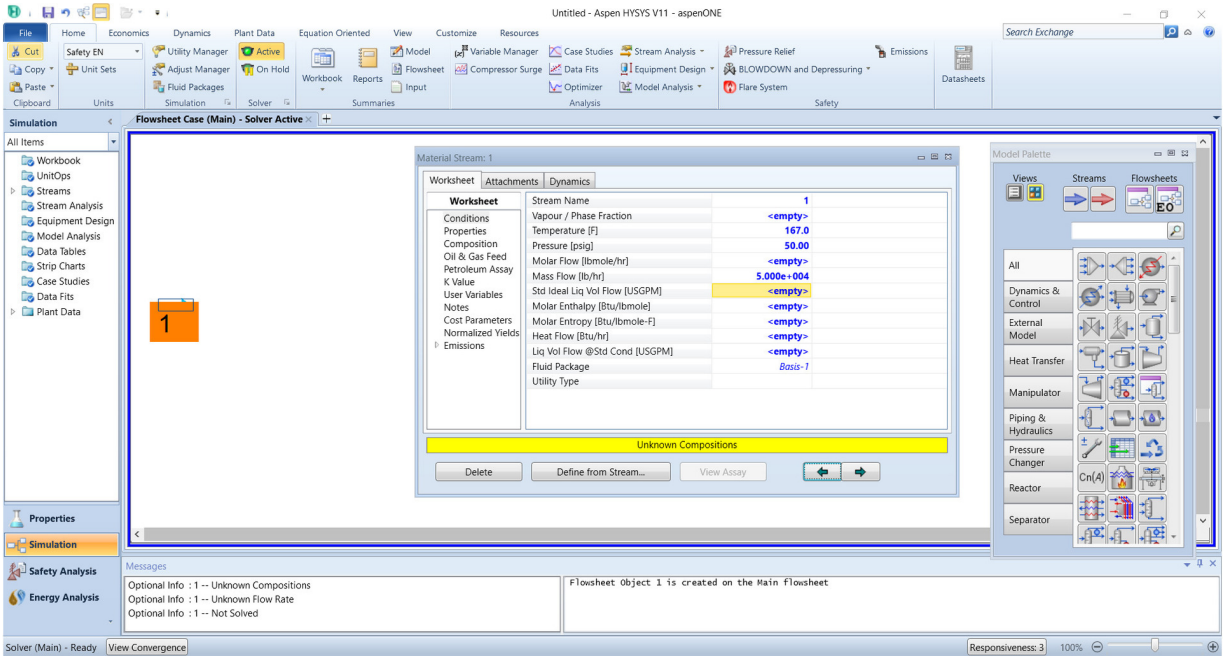
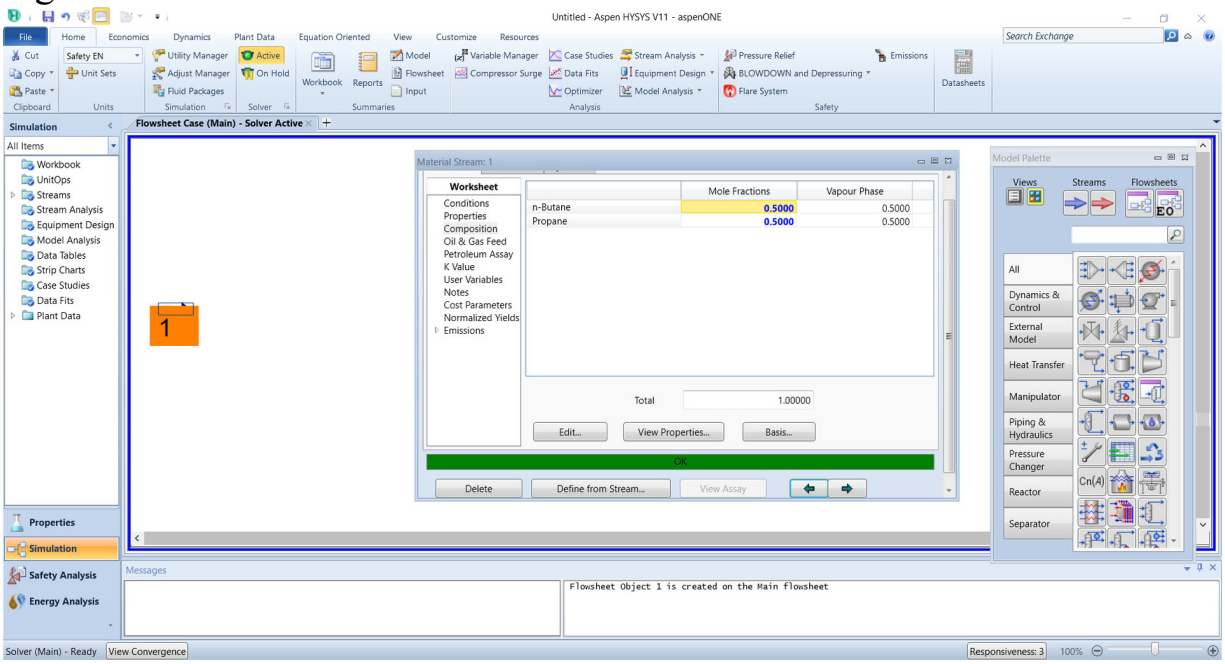
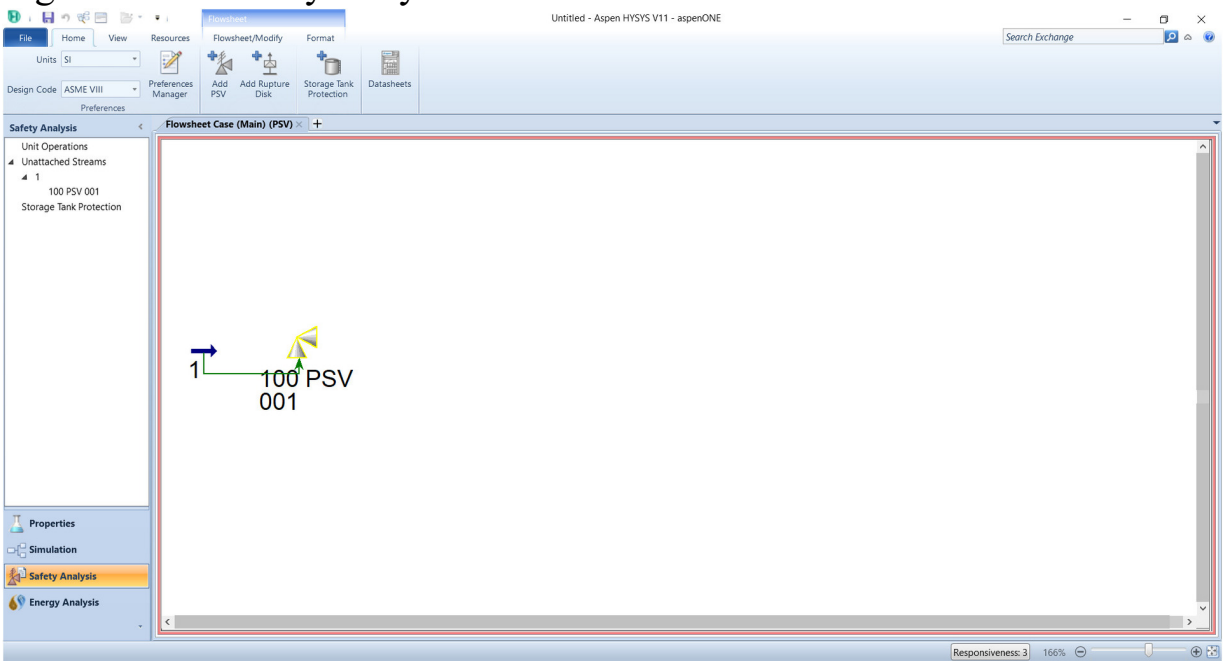


Figure 13.14: Material stream



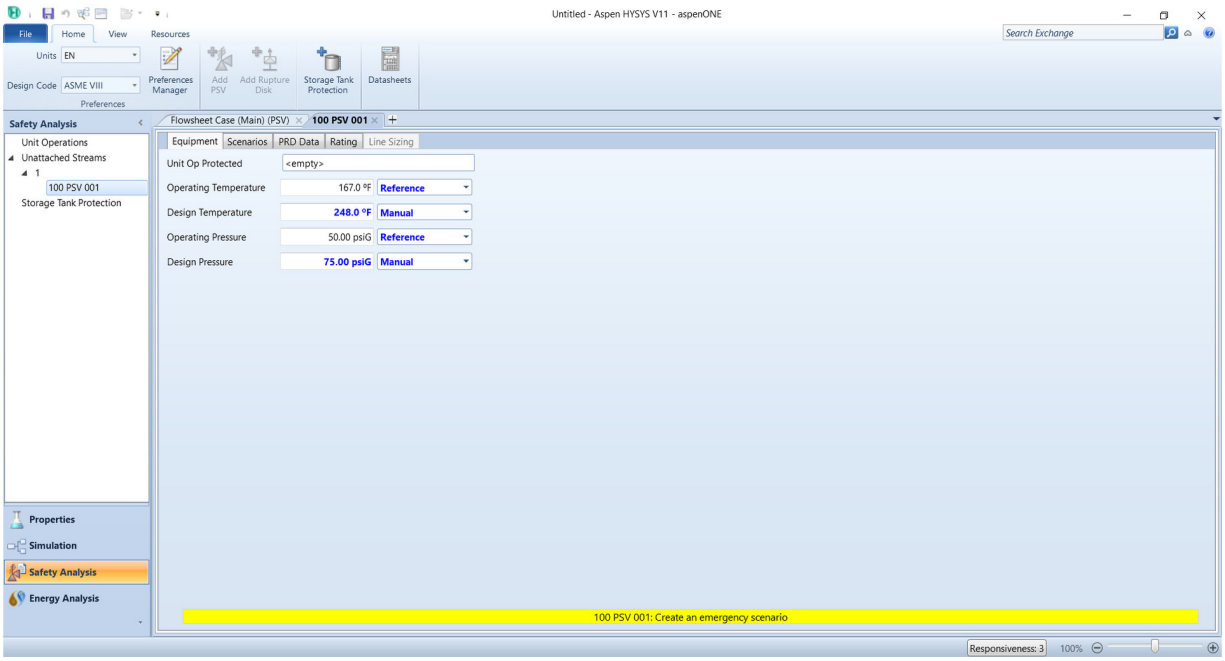
Next to enter “Safety Analysis” and “Add PSV” to the material stream “1”. It can be done by hovering the mouse over stream “1” and dragging the PSV out.

Figure 13.15: Safety analysis



Double click on PSV and provide operating/design temperature/pressure values as per the problem statement. “Reference” means the values will be referred from the converged simulation model and “Manual” means the values can be typed.

Figure 13.16: Equipment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Storage Tank Protection
Flowsheet Case (Main) (PSV)
100 PSV 001
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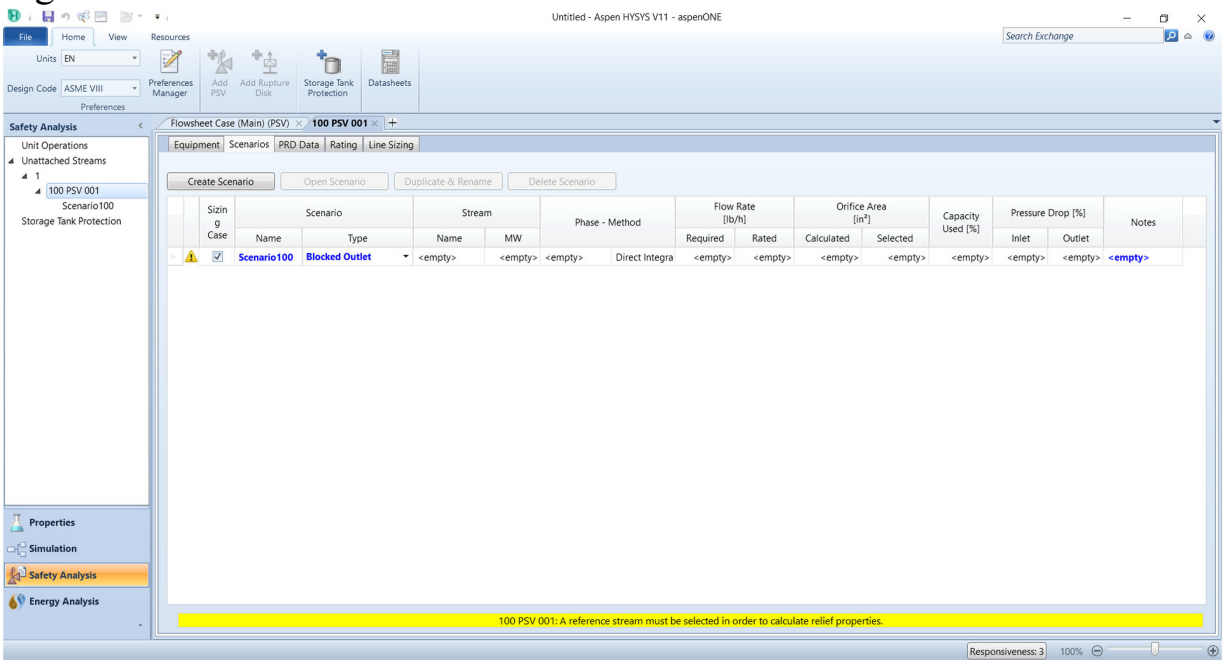
PRD Data
Rating
Line Sizing
Unit Op Protected
<empty>
Operating Temperature
167.0 °F
Reference
Design Temperature
248.0 °F
Manual
Operating Pressure
50.00 psiG
Reference
Design Pressure
75.00 psiG
Manual
100 PSV 001: Create an emergency scenario
Properties
Simulation
Safety Analysis
Energy Analysis
Responsiveness: 3
100%

Select the “Scenario” tab. One can have several scenarios created. But in this example, there is only one scenario, hence it will be ticked as “Sizing Case”. Scenario type selected is “Blocked outlet”, as the relief scenario is triggered by operational upset as per problem statement. Double click on “Scenario Name” to “Open Scenario” window.

Figure 13.17: Scenario



“Select” appropriate material stream. Provide “Relieving Temperature” using “Reference”. Relieving pressure shall be Set pressure + Accumulation. In this case 75 psig + 7.5 psi = 82.5 psig. One can view it by clicking on “Edit”. Total backpressure shall be 7.499 psig as super-imposed back pressure (maybe PSV is open to atmosphere) and built-up backpressure considered as 7.499 psi (< 10% of set pressure for conventional PSV).
From the Relieving Phase - Method drop-down list, you can select the calculation method used for the relieving device section. The Relieving Phase is determined based on a flash to the relief pressure and temperature. You can select the orifice sizing method. The following options are available:

Vapor: Used for vapor phase systems when the vapor compressibility factor is preferably within the range of $0.8<Z<1.1$. If used for a mixed-phase system, the vapor phase properties will be used for the sizing calculation.

Capacity-Certified Liquid: Method used for liquid relief through a valve with a certified liquid capacity. An assumption of incompressible liquid is used. For saturated liquids, a mixed-method can be used. If used for a mixed-phase system, the liquid phase properties will be used for the sizing calculation.

Non-Capacity-Certified Liquid: Method used for liquid relief through a valve not requiring capacity certification. Recommended for valves built for older API standards or valves that are not certified for liquid relief but may encounter scenarios in which they must pass an all-liquid stream.

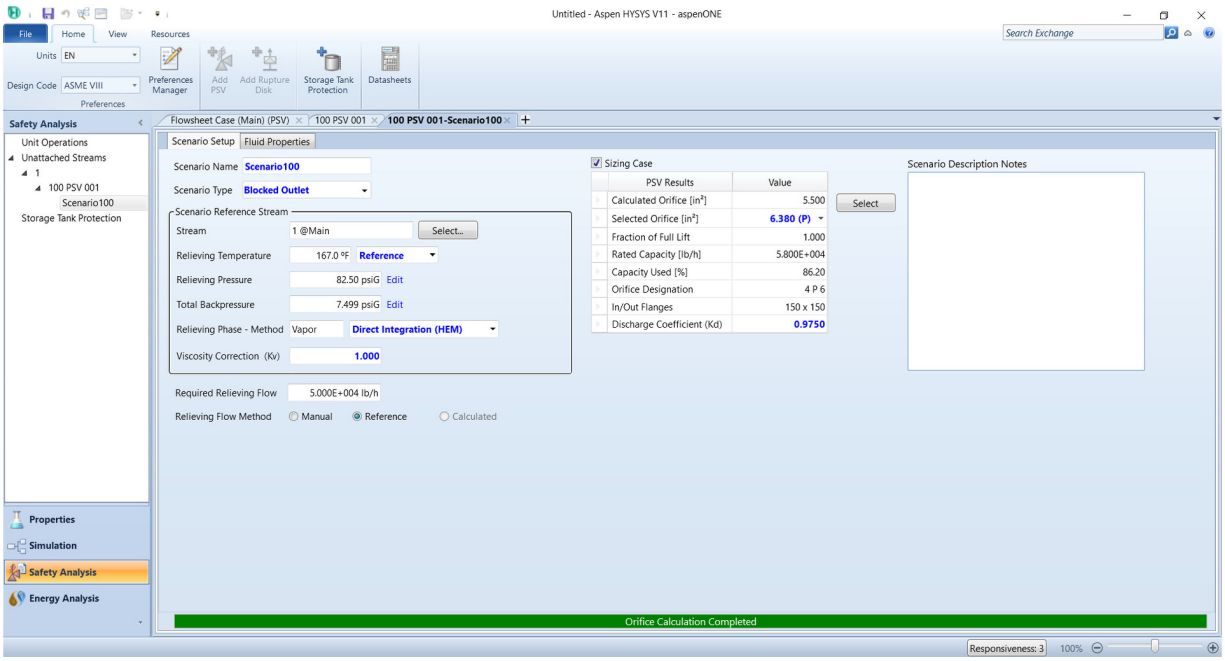
Steam: Method used for steam relief calculations at critical flow conditions. For steam at temperatures above 1200 ^{o} F, the vapor phase calculation method should be used instead.

Direct Integration (HEM): Sizing by the direct integration of the isentropic nozzle flow. General-purpose calculation method applicable to all single-phase and mixed-phase systems, for example, vapor cases where $Z<0.8$. This is the default sizing method for all cases. This is the recommended method for mixed-phase or supercritical fluids.
Notes:
For Fire scenarios using the Supercritical calculation method, make sure that the Direct Integration (HEM) method is selected.

For Fire scenarios using the Semi-Dynamic Flash calculation method, make sure that the Vapor or Steam method is selected.

“Required Relieving Flow” can be set “Manually” or by “Reference”.

Aspen HYSYS will calculate orifice size as per API 520 Part 1. By clicking “Select” it will choose appropriate Orifice size and inlet and outlet PSV flange size as per API 526. In this case, PSV designation is 4P6, meaning inlet flange size is 4”, outlet flange size is 6” and PSV is “P”. PSV discharge coefficient value should be referred from the vendor catalog. However, in the absence of the same, one can use preliminary values as 0.975 for vapor, 0.65 for liquid, and 0.85 for the two-phase system.
Figure 13.18: Scenario setup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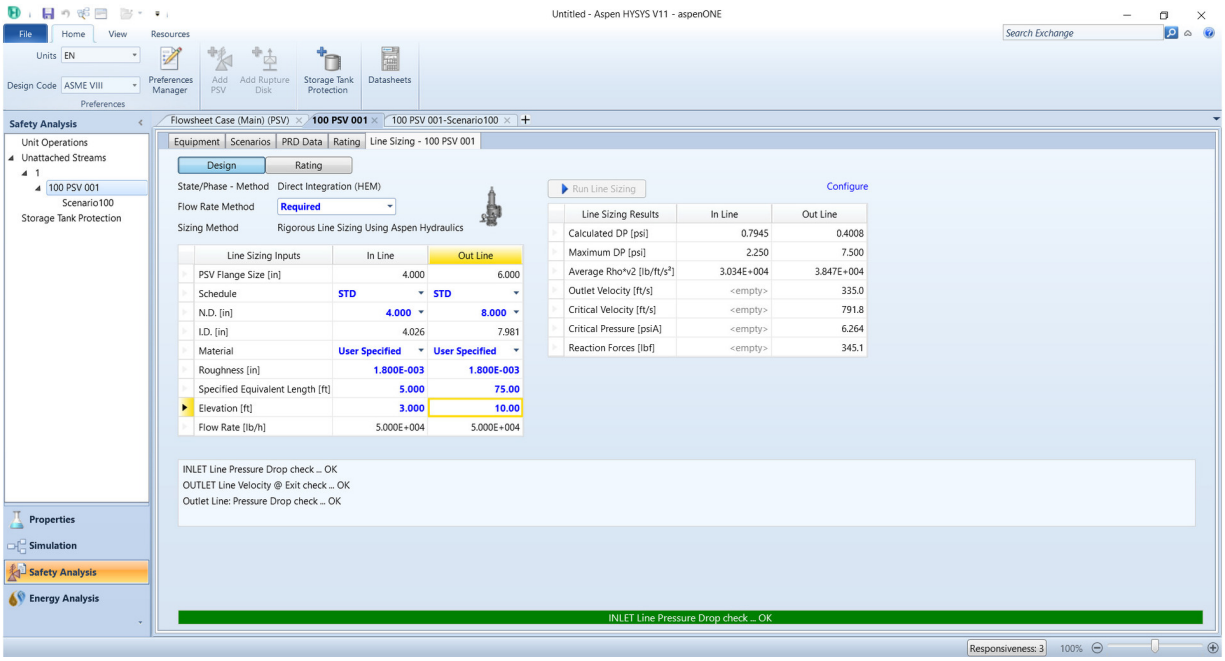
Scenario Setup
Fluid Properties
Scenario Name Scenario100
Scenario Type Blocked Outlet
Scenario Reference Stream
Stream 1 @Main
Select...
Relieving Temperature 167.0 °F Reference
Relieving Pressure 82.50 psiG Edit
Total Backpressure 7.499 psiG Edit
Relieving Phase - Method Vapor Direct Integration (HEM)
Viscosity Correction (Kv) 1.000
Required Relieving Flow 5.000E+004 lb/h
Relieving Flow Method Manual Reference Calculated
Sizing Case
PSV Results
Value
Calculated Orifice [in²] 5.500
Selected Orifice [in²] 6.380 (P)
Fraction of Full Lift 1.000
Rated Capacity [lb/h] 5.800E+004
Capacity Used [%] 86.20
Orifice Designation 4 P 6
In/Out Flanges 150 x 150
Discharge Coefficient (Kd) 0.9750
Select
Scenario Description Notes
Orifice Calculation Completed
Responsiveness: 3
100%

Now let's carry out line sizing of PSV inlet and outlet line. "Flow Rate Method" shall be "Required", as we want to keep the flow rate 50,000 lb/h. Pipe schedule of "Standard" is considered for inlet and outlet pipes. Line size is considered as 4" and 6" for inlet and outlet respectively. Material, length, and elevation data are also provided as shown in the below figure. Then click "Run Line Sizing". Aspen HYSYS checks if selected inlet line size is adequate w.r.to pressure drop criteria i.e. less than 3% of set pressure. In this case, set pressure is 75 psig, 3% of 75 is 2.25 psi. The calculated pressure drop in the inlet line is 0.7945 psi, hence selected inlet pipe size of 4" is adequate. It is always recommended to use PSV inlet pipe size equal or greater than PSV inlet flange size.

There are 2 checks carried out by HYSYS for the PSV outlet line (tailpipe). First, it calculates built-up backpressure, in this case, the value is 0.4006 psi, and then it compares with maximum allowable built-up backpressure. For conventional PSV it is 10% of set pressure (7.5 psi, in this example). Since the actual pressure drop is less than the allowable value, the selected tailpipe size is adequate. Additionally, it checks the Mach number and compares if the Mach number is less than 0.7 for tailpipe or not. Mach number is the ratio of fluid velocity to the sound velocity in that fluid. In this example, the fluid velocity is 334.9 ft/s and sound velocity is 791.7 ft/s. So Mach number is 0.42. Since it is less than 0.7, the tailpipe size is adequately selected in this example.

Figure 13.19: Line sizing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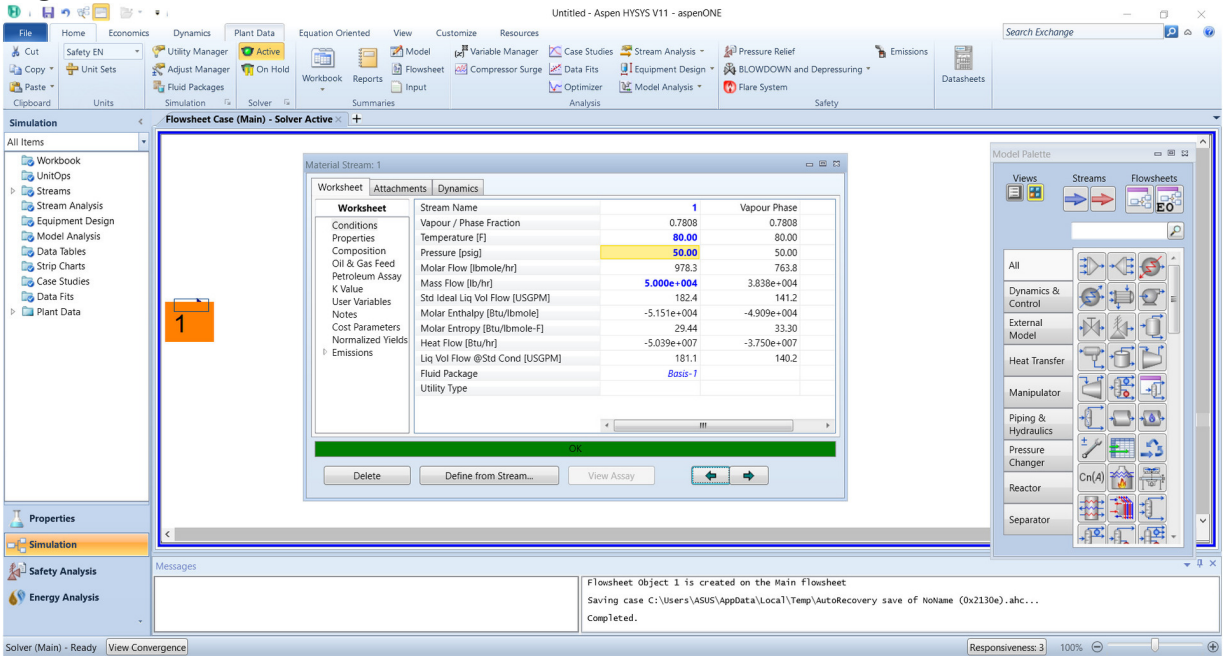
100 PSV 001-Scenario100
Equipment
Scenarios
PRD Data
Rating
Line Sizing - 100 PSV 001
Design
Rating
State/Phase - Method Direct Integration (HEM)
Flow Rate Method Required
Sizing Method Rigorous Line Sizing Using Aspen Hydraulics
Run Line Sizing
Configure
Line Sizing Inputs | In Line | Out Line
PSV Flange Size [in] | 4.000 | 6.000
Schedule | STD | STD
N.D. [in] | 4.000 | 8.000
I.D. [in] | 4.026 | 7.981
Material | User Specified | User Specified
Roughness [in] | 1.800E-003 | 1.800E-003
Specified Equivalent Length [ft] | 5.000 | 75.00
Elevation [ft] | 3.000 | 10.00
Flow Rate [lb/h] | 5.000E+004 | 5.000E+004
Line Sizing Results | In Line | Out Line
Calculated DP [psi] | 0.7945 | 0.4008
Maximum DP [psi] | 2.250 | 7.500
Average Rho*v2 [lb/ft/s²] | 3.034E+004 | 3.847E+004
Outlet Velocity [ft/s] | <empty> | 335.0
Critical Velocity [ft/s] | <empty> | 791.8
Critical Pressure [psiA] | <empty> | 6.264
Reaction Forces [lbf] | <empty> | 345.1
INLET Line Pressure Drop check ... OK
OUTLET Line Velocity @ Exit check ... OK
Outlet Line: Pressure Drop check ... OK
INLET Line Pressure Drop check ... OK
Responsiveness: 3
100%

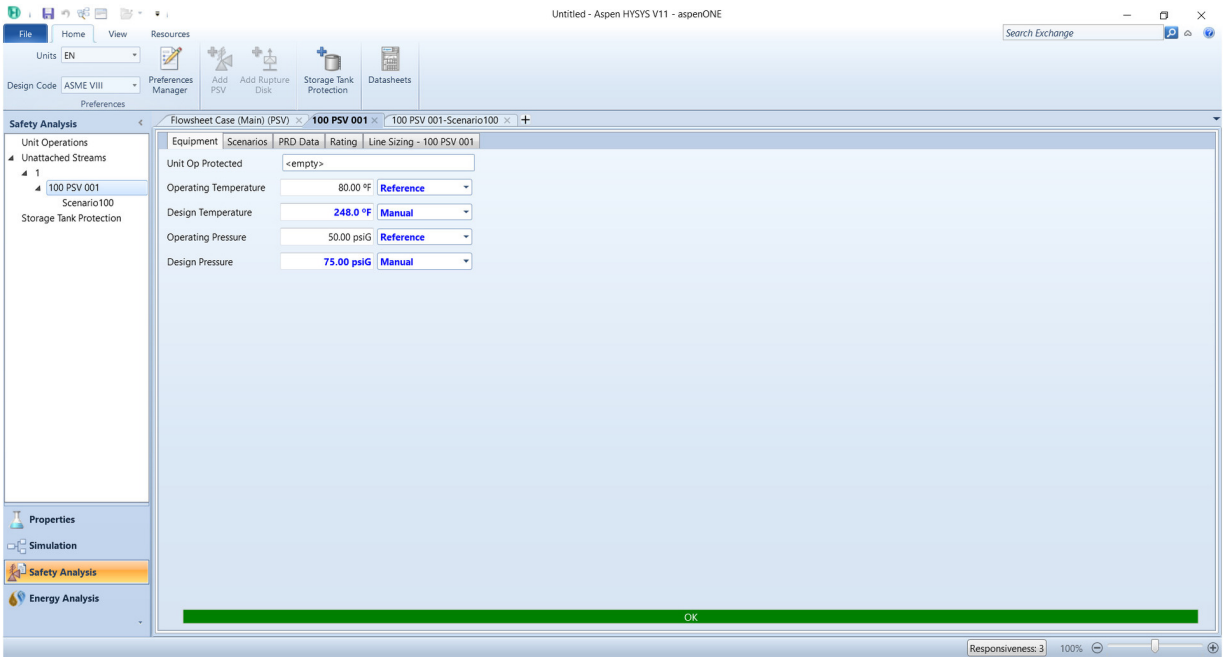
Let's now simulate PSV sizing for the fire case. The vessel is horizontal separator (3 ft diameter, 9 ft length), the liquid level is 2 ft. There is some change in temperature (80 ° F) inlet stream to the separator. This will make the vapor phase to 2 phase fluid. Separator is insulated vessel.

Figure 13.20: Material stream



Enter "Safety Analysis". Provide operating/design temperature/pressure as per the below figure. Select the "Scenario" tab.

Figure 13.21: Safety analysis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Scenario100
Storage Tank Protection
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100 PSV 001-Scenario100
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PRD Data
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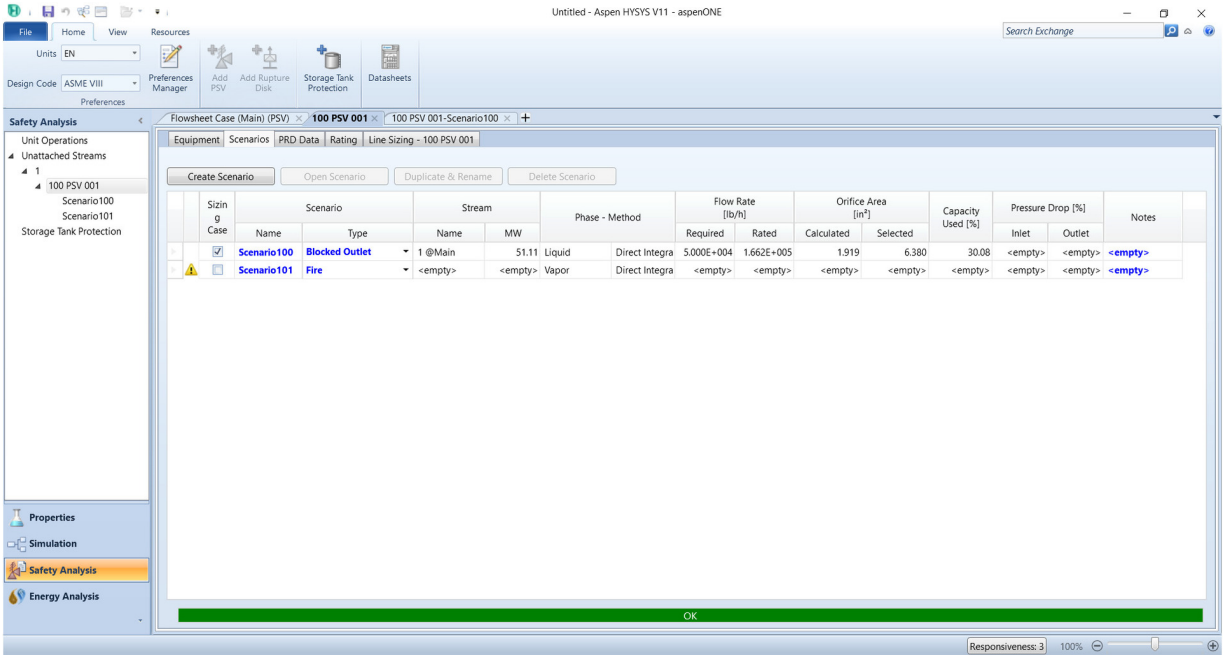
Line Sizing - 100 PSV 001
Unit Op Protected
<empty>
Operating Temperature
80.00 °F
Reference
Design Temperature
248.0 °F
Manual
Operating Pressure
50.00 psiG
Reference
Design Pressure
75.00 psiG
Manual
OK
Properties
Simulation
Safety Analysis
Energy Analysis
Responsiveness: 3
100%

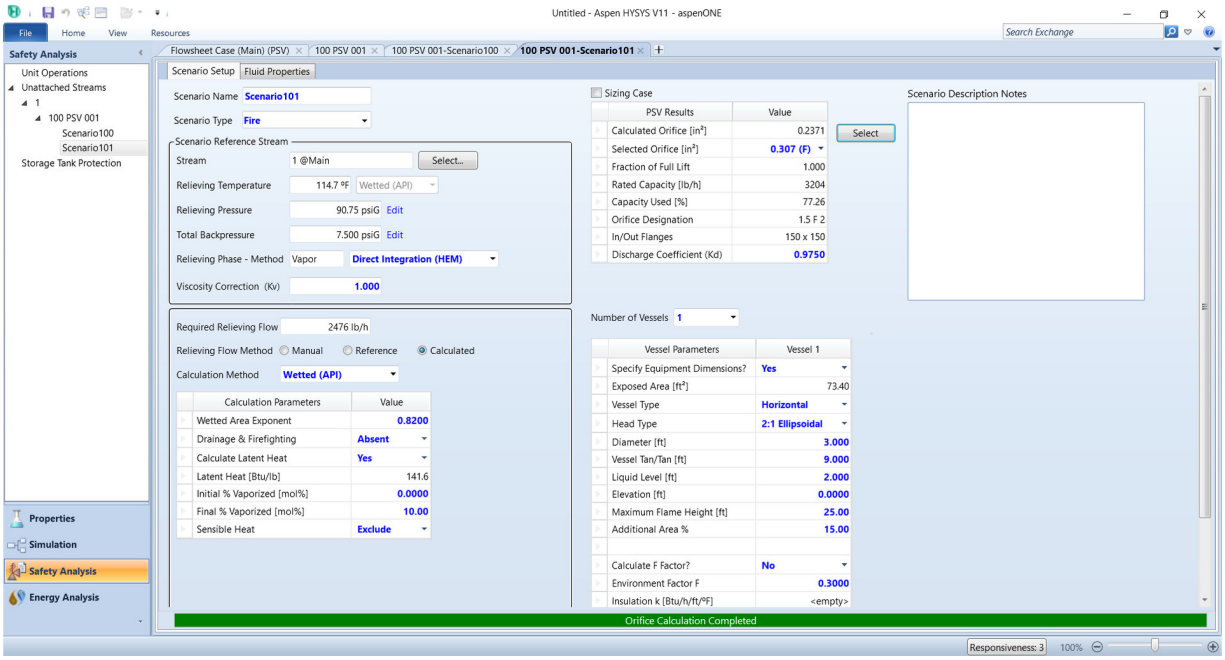
Select the “Scenario” tab and “Create Scenario”. This time we will choose the “Fire” scenario.

Figure 13.22: Scenario



Select the stream (should be the inlet stream to the separator). Relieving pressure includes 21% overpressure as it is a fire case. Provide vessel dimensions. As per API 521, F or environment factor is 1 for a bare vessel, for an insulated vessel, F is 0.3, for earth-covered or mounded vessel, F is 0.03. PSV size for the fire case calculated is 1.5F2.

Figure 13.23: Fire scenario



HYSYS calculates heat absorbed using the below formula.

$$Q = C \cdot F \cdot A_{ws}^{0.82}$$

Where;

Q is the total heat absorption (input) to the wetted surface, expressed in W (Btu/h);

C is a constant [= 43,200 in SI units (21,000 in USC units) when adequate drainage and firefighting Equipment exist];

[= 70,900 in SI units (34,500 in USC units) when adequate drainage and firefighting Equipment do not exist];

F is an environmental factor;

A_{ws} is the total wetted surface, expressed in m^2 (ft^2).

The heat absorbed by the wetted surface shall be used to increase the liquid temperature (sensible heat) and to vaporize liquid (latent heat). Accordingly, the vapor load is calculated by HYSYS. In this example, if sensible heat is excluded (which is negligible anyway), Q calculated to be 34500 x 0.3 x $73.4^{0.82}$ = 350600 Btu/h. Latent heat of vaporization is 141.6 Btu/lb. Hence vapor load will be 2476 lb/h. Once the vapor load is calculated, the same formulae from API 520 Part 1 can be used for PSV sizing.

Thermal Relief Valve (TRV)

Equipment and piping that can be blocked in and can be exposed to increased temperatures shall be protected against overpressure. TRVs may be installed with pocketed inlet and outlet lines where plugging and freezing within the inlet or outlet lines cannot occur or are prevented from occurring with tracing and insulation. TRVs shall discharge back into the process, the storage system, or a plant disposal system.

TRVs shall not discharge to an open drain unless the following are met:

- The discharged liquid cannot be accepted back into the process, storage, or plant disposal system.
- Any contained light hydrocarbon discharge is at a location where it will safely disperse.
- Stream does not contain any fluids defined as very toxic substances.

Identified liquid thermal expansion overpressure scenarios shall be mitigated with a pressure relief valve with an orifice area no smaller than a D. These D sized liquid relief valves are referred to as Thermal Expansion Relief Valves (TRVs). If the liquid thermal expansion

application is caused by heat gain from the environment, then relief load calculations and device sizing calculations are not required. The relief rate due to heat input is expected to be no larger than the rated flow of a nominally sized relief valve.

If the liquid thermal expansion application is caused by heat gain from heat tracing/jacketing, or the process, then the liquid thermal expansion relief load shall be calculated using the design duty of the equipment or heat gain from chemical reactions and the relief device sized. The relief device to mitigate liquid thermal relief should be located close to the process. The relief device does not need to be located at the highest point in the system.

Liquid thermal expansion overpressure protection shall be provided for piping that can be blocked in and contains more than 500 L (132 gals) of LPG or acute toxic liquids.

Per ASME B31.3, TRVs that protect only piping and piping components may be set at 120 % of the piping system design pressure. Depending on the design concept of a project (particularly about its required expandability, operability, and maintainability) may specify a lower allowable setpoint, e.g. 110 %.

Exercises

a. Mr. Aravinth is trying to size a PSV for the following scenario;
    - blocked outlet case,
    - 50,000 kg/h flow rate,
    - hydrocarbon vapor is a 50/50 (by mass) mixture of n-butane and propane,
    - relieving temperature, T, of 90 ^{o} C,
    - design pressure of the equipment is 5 barg,
    - backpressure of 0 psig
    - single PSV

Can you help him get the adequate PSV? Also, find out PSV inlet and outlet line sizes.

b. Mr. Pranit is trying to size PSVs for the following scenario;
    - blocked outlet case,
    - 50,000 kg/h flow rate,
    - hydrocarbon vapor is a 50/50 (by mass) mixture of n-butane and propane,
    - relieving temperature, T, of 90 ^{o} C,
    - design pressure of the equipment is 5 barg,
    - backpressure of 0 psig
    - 2 PSVs

Can you help him get the adequate PSV? Also, find out PSV inlet and outlet line sizes.

c. Mr. Omkar is trying to size a PSV for the following scenario;
    - fire case,
    - hydrocarbon vapor is a 50/50 (by mass) mixture of n-butane and propane,
    - separator size is 20 m^3 ,
    - design pressure of the equipment is 5 barg,
    - backpressure of 0 psig
    - single PSV

Can you help him get the adequate PSV? Also, find out PSV inlet and outlet line sizes.

d. Ms. Genevieve is trying to size a PSV for the following scenario;
    - Fire case,
    - operating temperature 10 ^{o} C,
    - hydrocarbon mixture of n-butane and propane with 5% vapor content (select operating pressure accordingly),
    - separator size is 20 m^3 ,
    - design pressure of the equipment is 5 barg,
    - backpressure of 0 psig
    - single PSV

Can you help her get the adequate PSV? Also, find out PSV inlet and outlet line sizes.

e. Mr. Sanchit is trying to size a PSV for the following scenario;
    - Fire case,
    - operating temperature 10 ^{o} C,
    - hydrocarbon mixture of n-butane and propane with 5% vapor content (select operating pressure accordingly),
    - separator size is 200 m^3 ,
    - design pressure of the equipment is 5 barg,
    - backpressure of 0 psig
    - single PSV

Can you help him get the adequate PSV? Also, find out PSV inlet and outlet line sizes.

f. Mr. Vivek is trying to size a rupture disk for the following scenario;
    - Blocked outlet case,
    - 50,000 kg/h flow rate,
    - hydrocarbon vapor is a 50/50 (by mass) mixture of n-butane and propane,
    - relieving temperature, T, of 90 ^{o} C,
    - design pressure of the equipment is 5 barg,
    - backpressure of 0 psig

Can you help him get the adequate rupture disk size?

g. Mr. Ashutosh is trying to size a TRV for the following scenario;
    - cross country (Qatar to Saudi Arabia) uninsulated pipeline,
    - Above ground pipeline (6", Sch. 40, 1000 m)
    - Fluid is Benzene,
    - Initial pressure 2 barg,
    - design pressure of the pipe 10 barg,

Can you help him get the adequate TRV size? Also, find out TRV inlet and outlet line sizes.

CHAPTER 14: PROCESS PLANT DESIGN USING ASPEN HYSYS

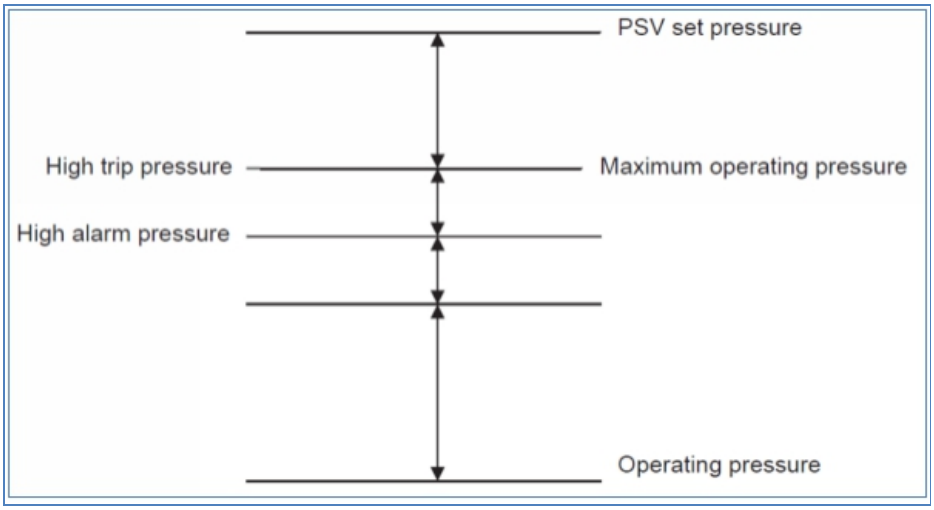
Process plant design is the combination of theoretical aspects and practical or experience-based aspects.
A process engineer must be able to consider appropriate numbers using the available standards and experience on similar projects.

Maximum design pressure
Below table will come handy to arrive at design pressure value of pressurized system;

Table 14.1: Design pressure guideline

Maximum operating pressure, barg	**Design pressure, barg**
0 to 35	Maximum operating pressure + 3.5 bar
35 to 70	Maximum operating pressure + 10%
70 to 200	Maximum operating pressure + 8.5%, but between (7 to 10 bar)
200 +	Maximum operating pressure + 5%

Figure 14.1: Pressure terms



Atmospheric tanks shall as a minimum be designed to be liquid filled to the highest point of the overflow line and with an overpressure of 0.07 bar.
For flare knock out drums, it is acceptable that the design pressure is equal to the maximum operating pressure.

Minimum design pressure
For equipment where cooling or condensing vapors, drainage or pump out may lead to less than atmospheric pressure, the equipment shall be designed for full vacuum or protected by vacuum relief.

Maximum design temperature
Where the maximum operating temperature can be determined accurately, this temperature should be used as maximum design temperature, without adding a safety margin.
Where the maximum operating temperature cannot be calculated accurately, the maximum design temperature should be determined by adding 30 ° C to the normal operating temperature.

Minimum design temperature
The minimum design temperature determines the requirements to the low-temperature characteristics of the material, and shall be the most stringent of the following:

- minimum operating temperature with a margin of 5 ° C.
- minimum ambient temperature
- minimum temperature occurring during depressurizing with a margin of 5 ° C

Line sizing criteria
Line sizing is carried out mainly w.r.to velocity criteria.

- Centrifugal pump suction line typical velocity is 1 m/s and discharge velocity is 2.7 m/s.

- Pressurized gas velocity should be comfortably less than 18 m/s.
- Gravity line velocity is kept around 1 m/s.
- Two-phase line sizing is carried out keeping fluid velocity comfortably less than erosion velocity.
- PSV inlet line pressure drop should be less than 3% of set pressure.
- PSV tailpipe or Depressurization line Mach number should not exceed 0.7.
- Flare header Mach number should be less than 0.5.
- Pressure drop in vent line for atmospheric tank should be less than 0.07 bar.

Separator sizing

Oil retention time shall depend on ° API of oil as per API 12J. If ° API is more than 35, then retention time can be considered as 1 minute. If it is between 20 to 30 ° API, the retention time may be taken as 1 to 2 minutes and if ° API is less than 20, then retention time may be considered as 2 to 4 minutes.

For mixed-phase feed with less vapor content, horizontal separator may be suggested and vice versa.

Typical L/D ratio for horizontal separator is 2.5 to 5 and H/D ratio for vertical separator is 2 to 4.

Atmospheric tank design

For atmospheric tanks containing flammable liquids, the purge/blanketing gas flow rate shall be at least 5% higher than maximum pumping capacity.

In some cases, flame arrestors are recommended for use on atmospheric tanks. The use of a flame arrestor within the tank's relief path introduces the risk of overpressure due to the flame arrestor clogging/fouling. Flame arrestors shall be located for easy access and included in a rigorous maintenance program. Continuous monitoring of differential pressure across the flame arrestor is advised

if the process fluid is known to contain particulates or substances which may block the flame arrestor. The materials in the vent pipe should be corrosion resistant to reduce the risk of plugging the flame arrestor.

Heat exchanger design

For all heat exchangers, both sides shall have the same maximum design temperature determined by the hottest of the fluids on either side.

The maximum continuous duty will determine the heat exchanger area.

Fouling or extra heat exchanger area should be added as a design margin.

Operational drain, vent and flushing requirements

All equipment and piping shall be provided with high point vents and low point drains within isolation valves isolating equipment or process sections. All such vents and drains shall be fitted with a valve and blind flange. For piping and headers, low point drain and high point vent shall, in general, be provided.

Steam-out and utility connections shall be provided and located to ensure efficient flushing and cleaning required for inspection and maintenance.

Where provisions are made for chemical cleaning of heat exchangers with the tube bundle in place, blind flange connections shall be provided for chemical hose attachments. The connections shall be minimum DN 80 (3 in), but not exceeding line size, and shall be located between the exchanger nozzles and the block valves.

Plant design problem

So far, we have simulated control valve, PSV, line, orifice, separator, and pump. We are now going to use this skill to design a real plant.

Simulation

Problem statement is given below;

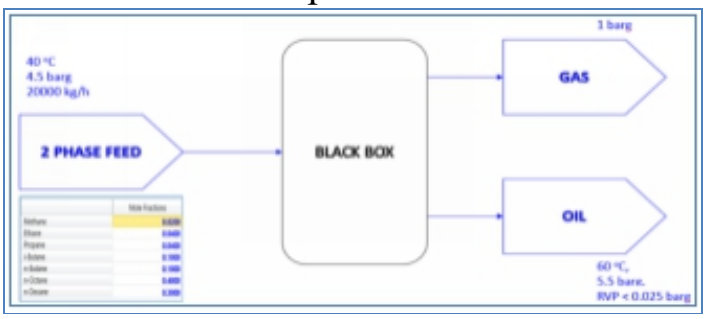
We need to design a separation system to separate gas from oil in a 2 phase fluid, that is available at 40 ° C, 4.5 barg pressure, and 20000 kg/h. Mole fraction of feed is provided in the below table.

Table 14.2: Feed composition

Component	**Mole fraction**
Methane	0.02
Ethane	0.04
Propane	0.04
i-Butane	0.10
n-Butane	0.10
n-Octane	0.40
n-Decane	0.30

The gaseous product shall be connected to the existing LP flare header at 1 barg pressure. Oil shall be sent to the existing header at 60 ° C and 5.5 barg pressure. RVP of the oil shall be less than 0.025 barg.

Figure 14.2: Schematic of problem



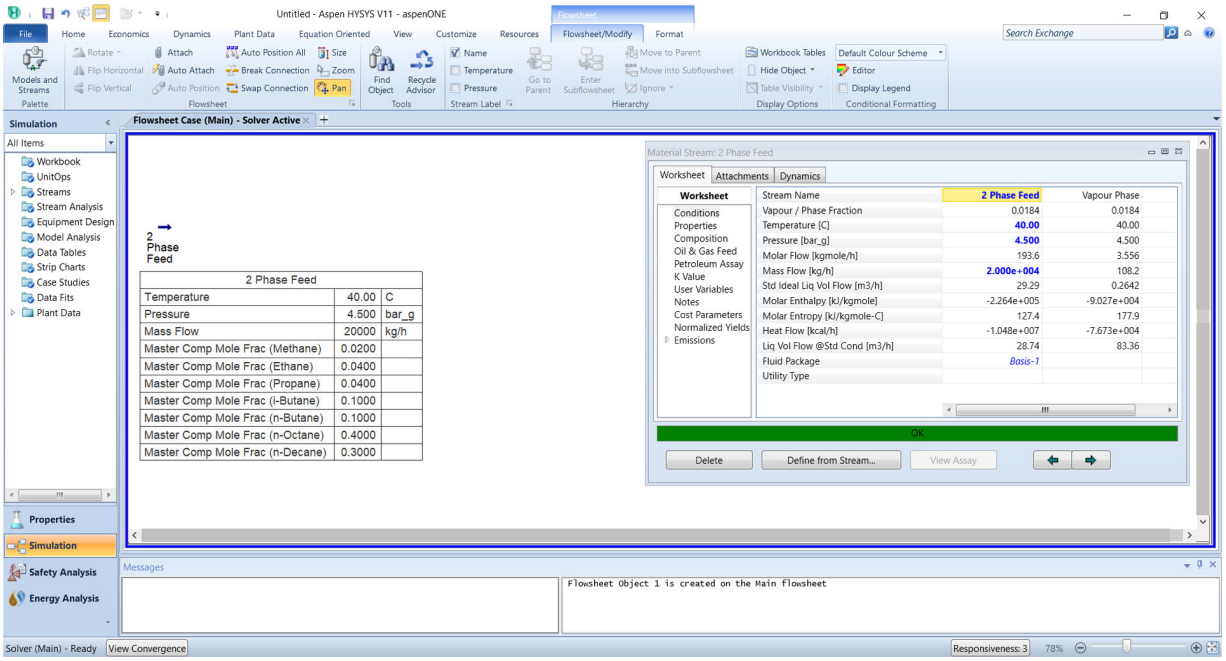
The black box signifies all the unit operations/calculations/sizing that is needed to achieve the project objectives.

What does the black box contain?

- Separator sizing
- Heater sizing
- Pump sizing

- Line sizing
- LCV sizing
- PCV sizing
- FCV sizing
- Flow orifice sizing
- PSV sizing

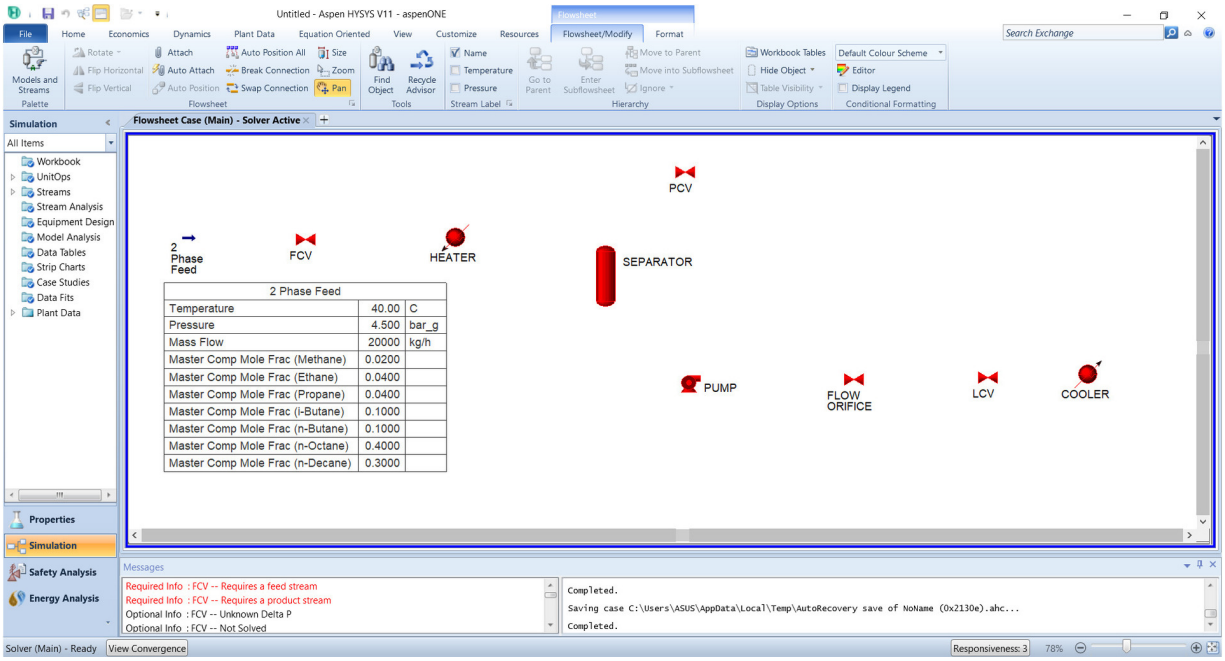
Let's simulate.
First, get the components added and the fluid package selected shall be Peng Robinson.
The feed stream is converged.
Figure 14.3: Material stream



2 Phase Feed		
Temperature	40.00	C
Pressure	4.500	bar_g
Mass Flow	20000	kg/h
Master Comp Mole Frac (Methane)	0.0200	
Master Comp Mole Frac (Ethane)	0.0400	
Master Comp Mole Frac (Propane)	0.0400	
Master Comp Mole Frac (I-Butane)	0.1000	
Master Comp Mole Frac (n-Butane)	0.1000	
Master Comp Mole Frac (n-Octane)	0.4000	
Master Comp Mole Frac (n-Decane)	0.3000	

Place all the equipment in order.

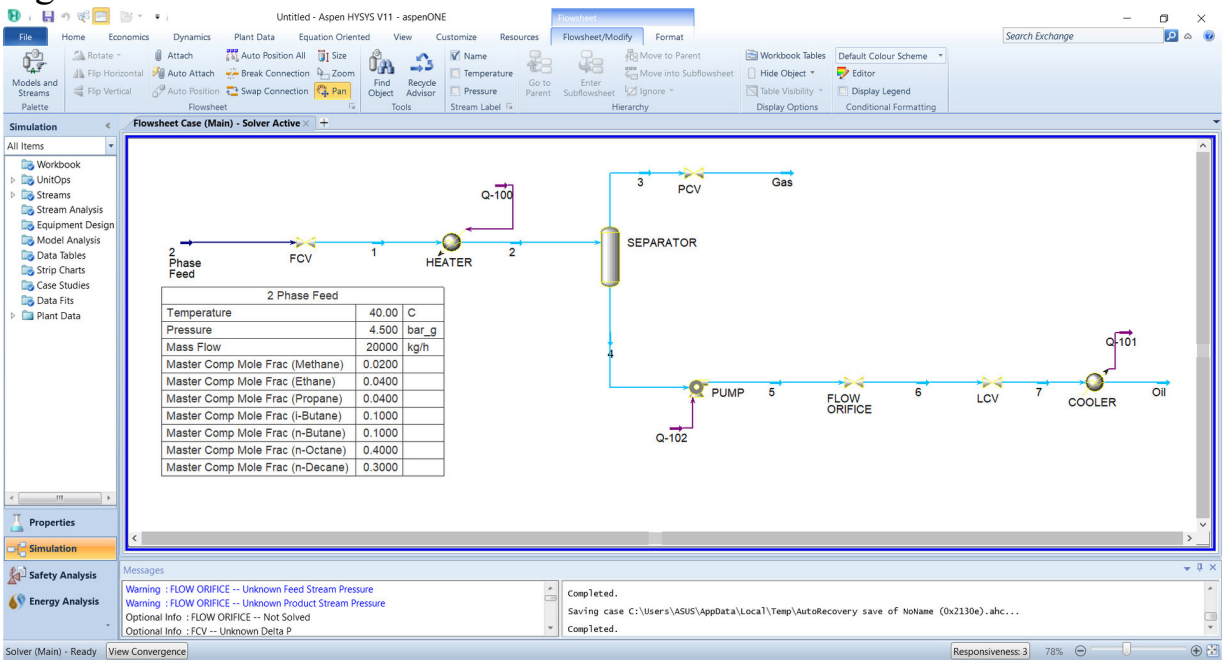
Figure 14.4: Process scheme



2 Phase Feed		
Temperature	40.00	C
Pressure	4.500	bar_g
Mass Flow	20000	kg/h
Master Comp Mole Frac (Methane)	0.0200	
Master Comp Mole Frac (Ethane)	0.0400	
Master Comp Mole Frac (Propane)	0.0400	
Master Comp Mole Frac (i-Butane)	0.1000	
Master Comp Mole Frac (n-Butane)	0.1000	
Master Comp Mole Frac (n-Octane)	0.4000	
Master Comp Mole Frac (n-Decane)	0.3000	

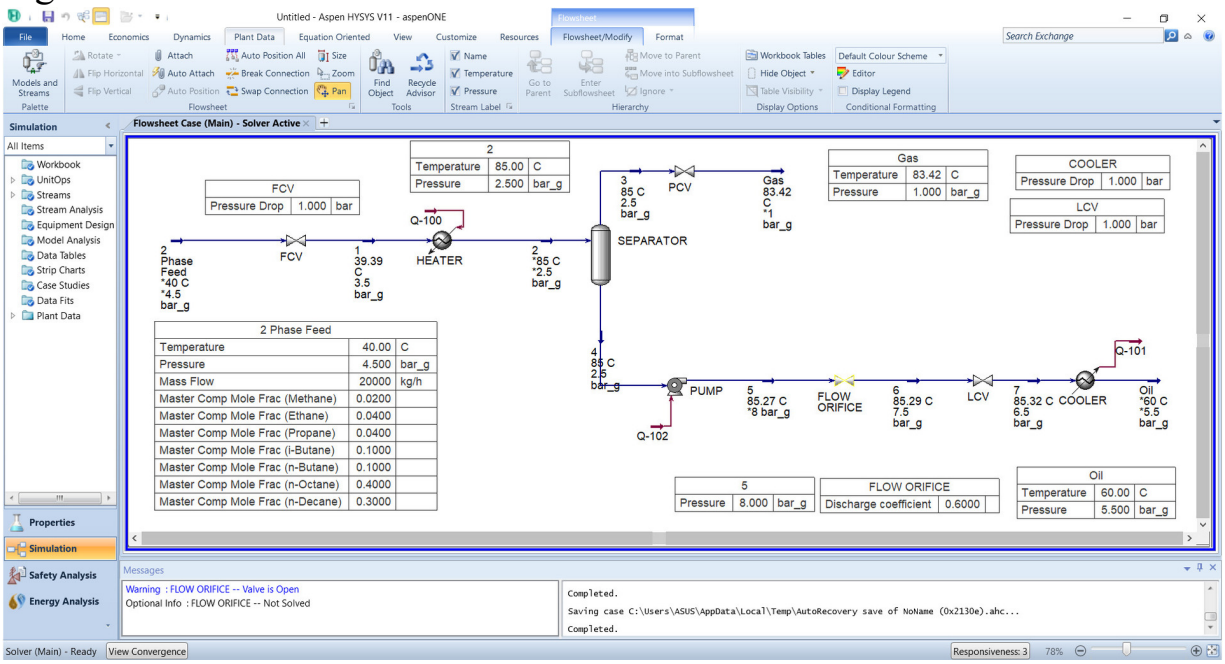
Connect all the equipment in order.

Figure 14.5: Process scheme



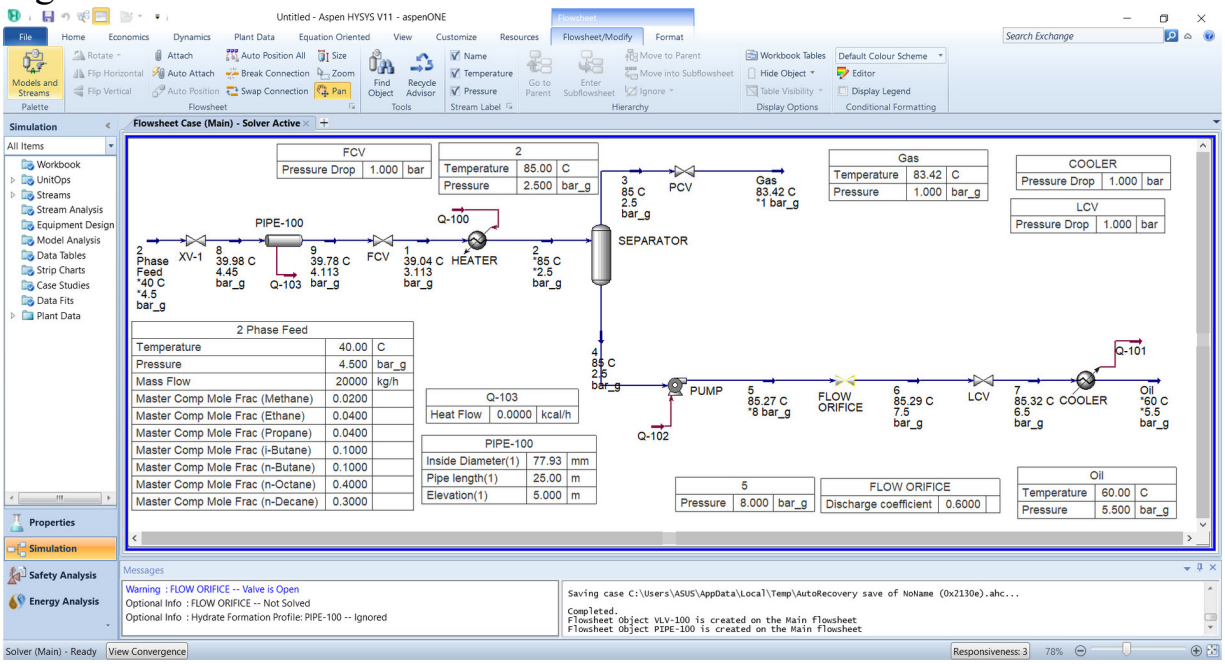
Provide all the available process parameters and get the model converged.

Figure 14.6: Process flowsheet



Let’s also add an ON/OFF valve and a pipe segment upstream of FCV.

Figure 14.7: Process scheme

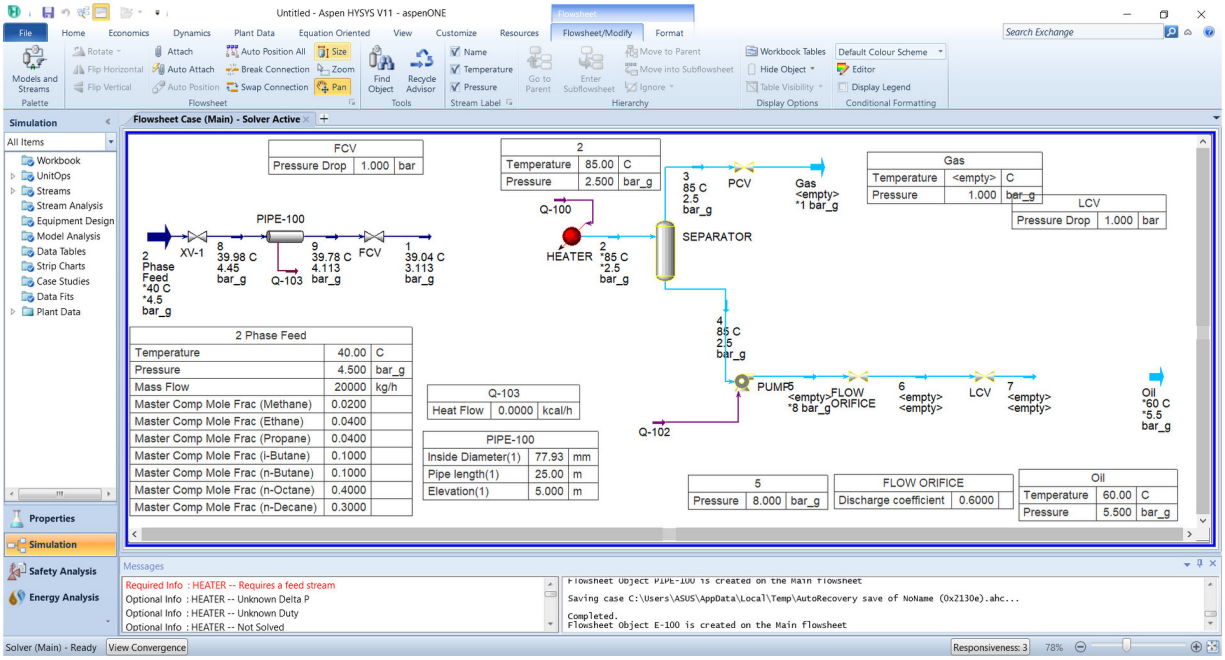


Is this the best design possible?

- Heater duty = 697 kW
- Cooler duty = 307 kW

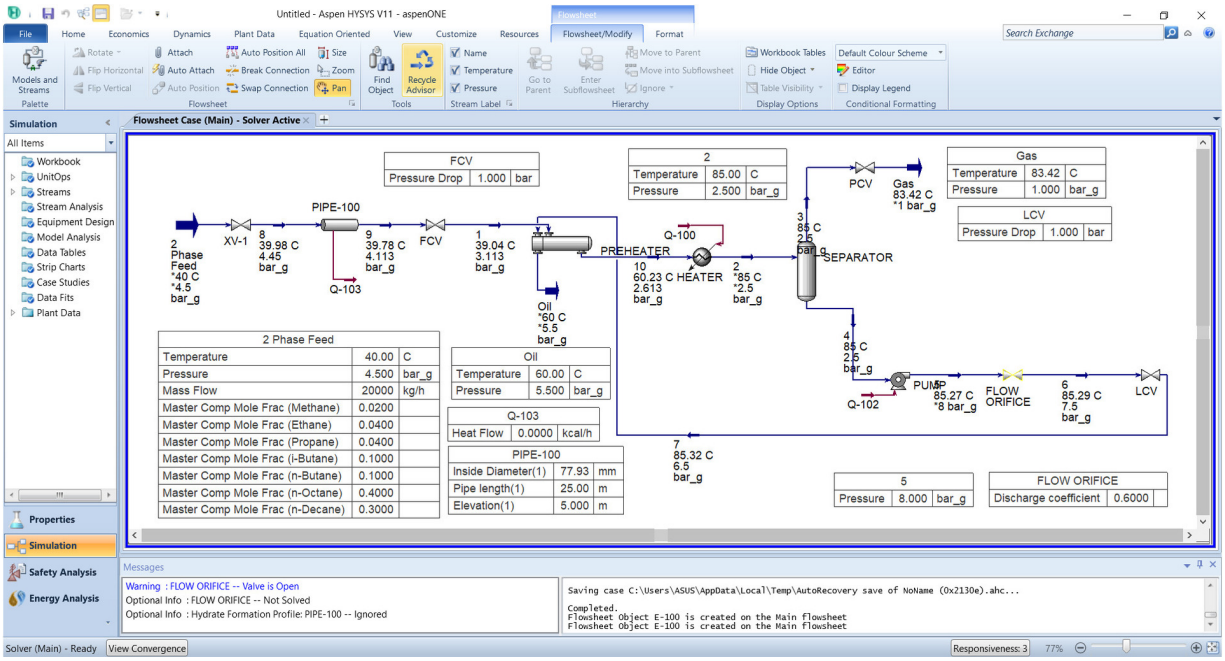
Can we optimize this design?
Remove the cooler and add a pre-heater (heat exchanger) downstream of FCV. Feed shall be pre-heated first by hot oil from separator before entering the heater.

Figure 14.8: Process scheme



Feed shall the shell side fluid and hot oil shall be the tube side fluid in the pre-heater. Pressure drop across pre-heater (E-100) and Heater is considered at 0.5 bar each.

Figure 14.9: Process scheme


Untitled - Aspen HYSYS V11 - aspenONE
Flowsheet Case (Main) - Solver Active
FCV
Pressure Drop 1.000 bar
PIPE-100
XV-1
Q-103
FCV
PREHEATER
Q-100
HEATER
SEPARATOR
PCV
Gas
83.42 C
*1 bar_g
Oil
*60 C
*5.5
bar_g
Q-102
PUMP
FLOW ORIFICE
LCV
2 Phase Feed
Temperature 40.00 C
Pressure 4.500 bar_g
Mass Flow 20000 kg/h
Master Comp Mole Frac (Methane) 0.0200
Master Comp Mole Frac (Ethane) 0.0400
Master Comp Mole Frac (Propane) 0.0400
Master Comp Mole Frac (I-Butane) 0.1000
Master Comp Mole Frac (n-Butane) 0.1000
Master Comp Mole Frac (n-Octane) 0.4000
Master Comp Mole Frac (n-Decane) 0.3000
Oil
Temperature 60.00 C
Pressure 5.500 bar_g
Q-103
Heat Flow 0.0000 kcal/h
PIPE-100
Inside Diameter(1) 77.93 mm
Pipe length(1) 25.00 m
Elevation(1) 5.000 m
2
Temperature 85.00 C
Pressure 2.500 bar_g
Gas
Temperature 83.42 C
Pressure 1.000 bar_g
LCV
Pressure Drop 1.000 bar
5
Pressure 8.000 bar_g
FLOW ORIFICE
Discharge coefficient 0.6000
Warning : FLOW ORIFICE -- Valve is Open
Optional Info : FLOW ORIFICE -- Not Solved
Optional Info : Hydrate Formation Profile: PIPE-100 -- Ignored
Saving case C:\Users\ASUS\AppData\Local\Temp\AutoRecovery save of NoName (0x2130e).ahc...
Completed.
Flowsheet Object E-100 is created on the Main Flowsheet
Flowsheet Object E-100 is created on the Main Flowsheet
Solver (Main) - Ready
View Convergence

Revised heat duty is given below;

- Heater duty = 390 kW (from 697 kW, 40% savings)
- Cooler duty = 0 kW (from 307 kW, 100% savings)

Let's design.

Separator sizing (as per standard API 12J)

Figure 14.10: Separator sizing

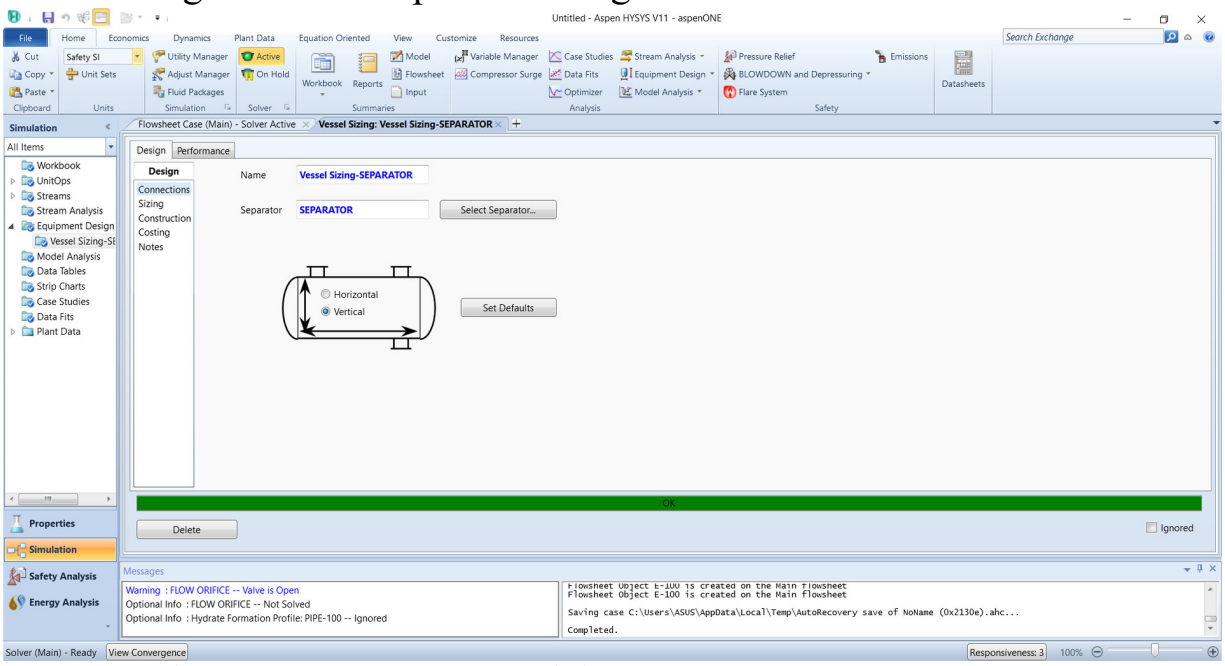


Figure 14.11: Separator sizing

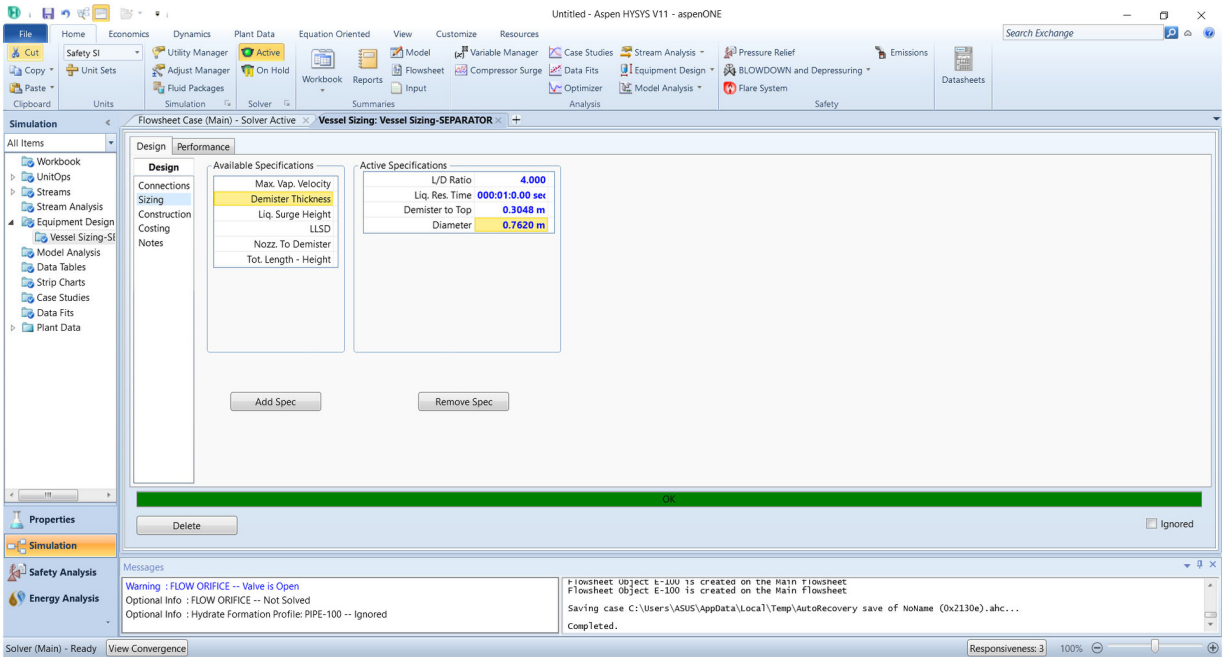
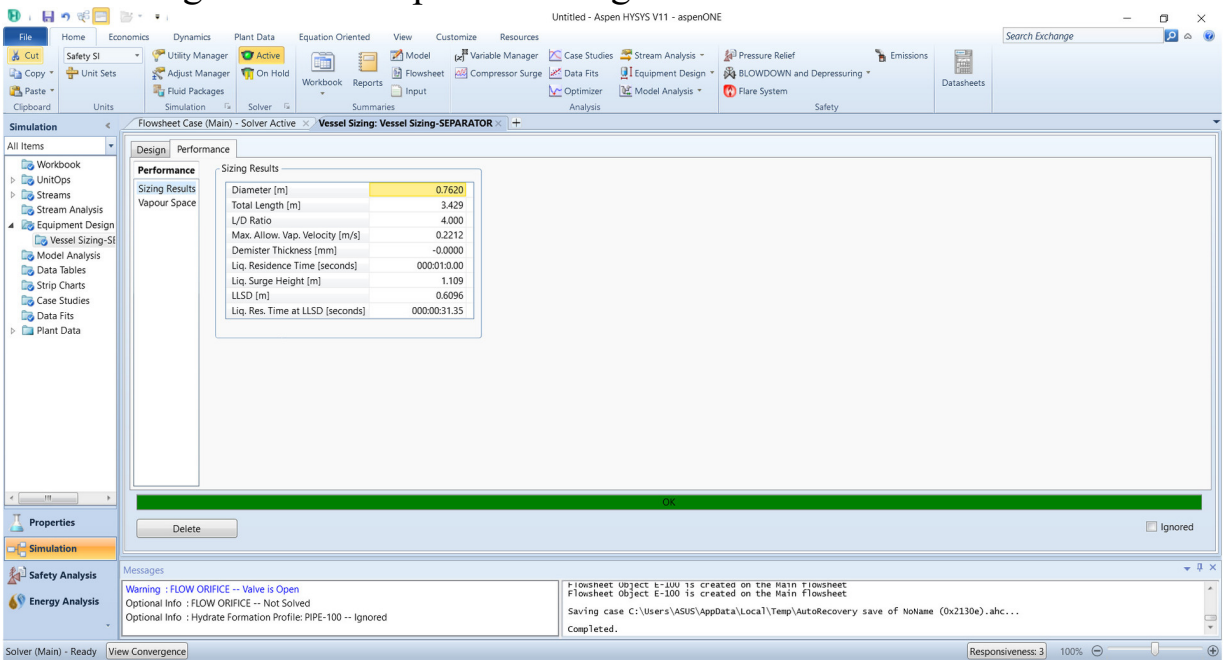

Untitled - Aspen HYSYS V11 - aspenONE
File
Home
Economics
Dynamics
Plant Data
Equation Oriented
View
Customize
Resources
Search Exchange
Cut
Copy
Paste
Clipboard
Safety SI
Unit Sets
Units
Utility Manager
Adjust Manager
Fluid Packages
Simulation
Active
On Hold
Solver
Workbook
Reports
Model
Flowsheet
Input
Summaries
Variable Manager
Compressor Surge
Case Studies
Data Fits
Optimizer
Stream Analysis
Equipment Design
Model Analysis
Analysis
Pressure Relief
BLOWDOWN and Depressuring
Flare System
Emissions
Safety
Datasheets
Simulation
All Items
Workbook
UnitOps
Streams
Stream Analysis
Equipment Design
Vessel Sizing-SI
Model Analysis
Data Tables
Strip Charts
Case Studies
Data Fits
Plant Data
Flowsheet Case (Main) - Solver Active
Vessel Sizing: Vessel Sizing-SEPARATOR
Design
Performance
Design
Connections
Sizing
Construction
Costing
Notes
Available Specifications
Max. Vap. Velocity
Demister Thickness
Liq. Surge Height
LLSD
Nozz. To Demister
Tot. Length - Height
Active Specifications
L/D Ratio 4.000
Liq. Res. Time 000:01:0.00 sec
Demister to Top 0.3048 m
Diameter 0.7620 m
Add Spec
Remove Spec
OK
Delete
Ignored
Properties
Simulation
Safety Analysis
Energy Analysis
Messages
Warning : FLOW ORIFICE -- Valve is Open
Optional Info : FLOW ORIFICE -- Not Solved
Optional Info : Hydrate Formation Profile: PIPE-100 -- Ignored
Flowsheet Object E-100 is created on the Main Flowsheet
Flowsheet Object E-100 is created on the Main Flowsheet
Saving case C:\Users\ASUS\AppData\Local\Temp\AutoRecovery save of NoName (0x2130e).ahc...
Completed.
Solver (Main) - Ready
View Convergence
Responsiveness: 3
100%

Figure 14.12: Separator sizing



- Orientation = Vertical is selected because the feed has more of vapors.
- Diameter = 2.5 ft
- Height = 10 ft (T/T)
- Liquid % level = 35%

Heat Exchanger (E-100) sizing using Aspen EDR

Figure 14.13: Heat exchanger sizing

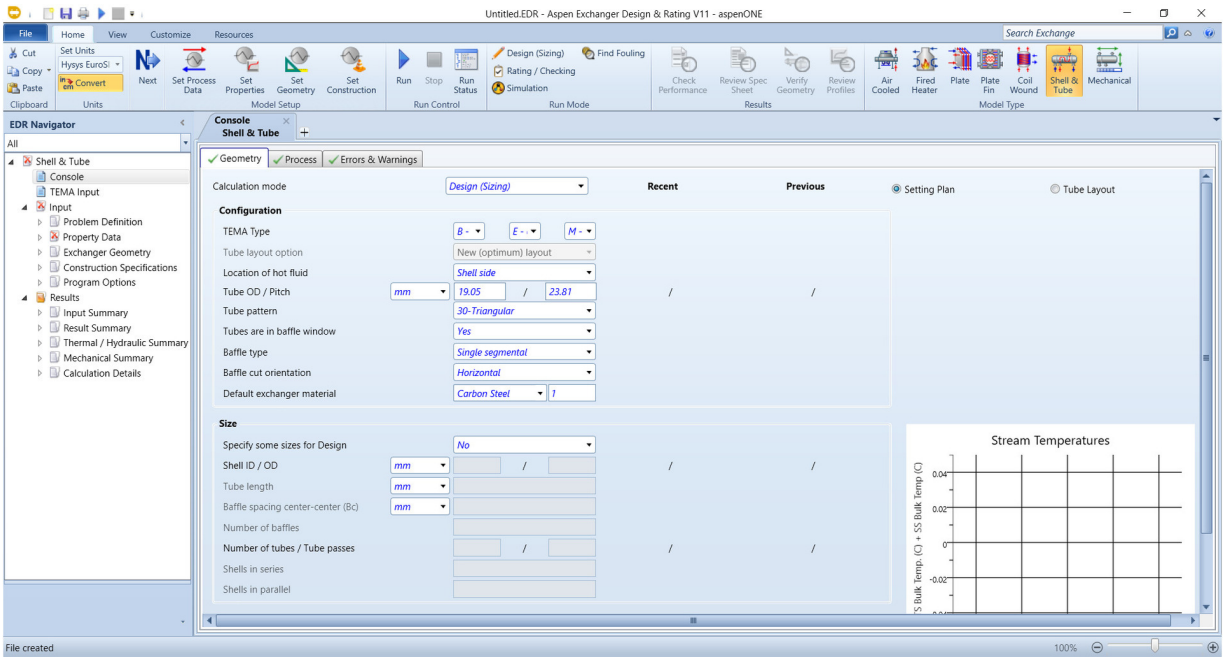


Figure 14.14: Heat exchanger sizing

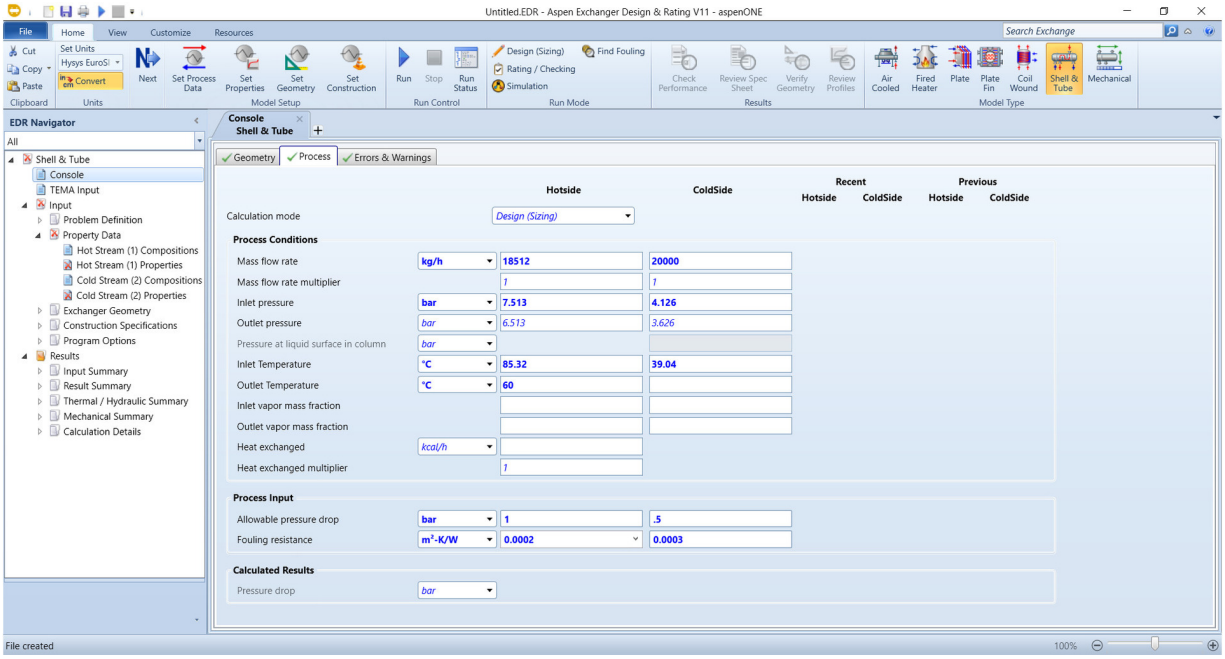


Figure 14.15: Heat exchanger sizing

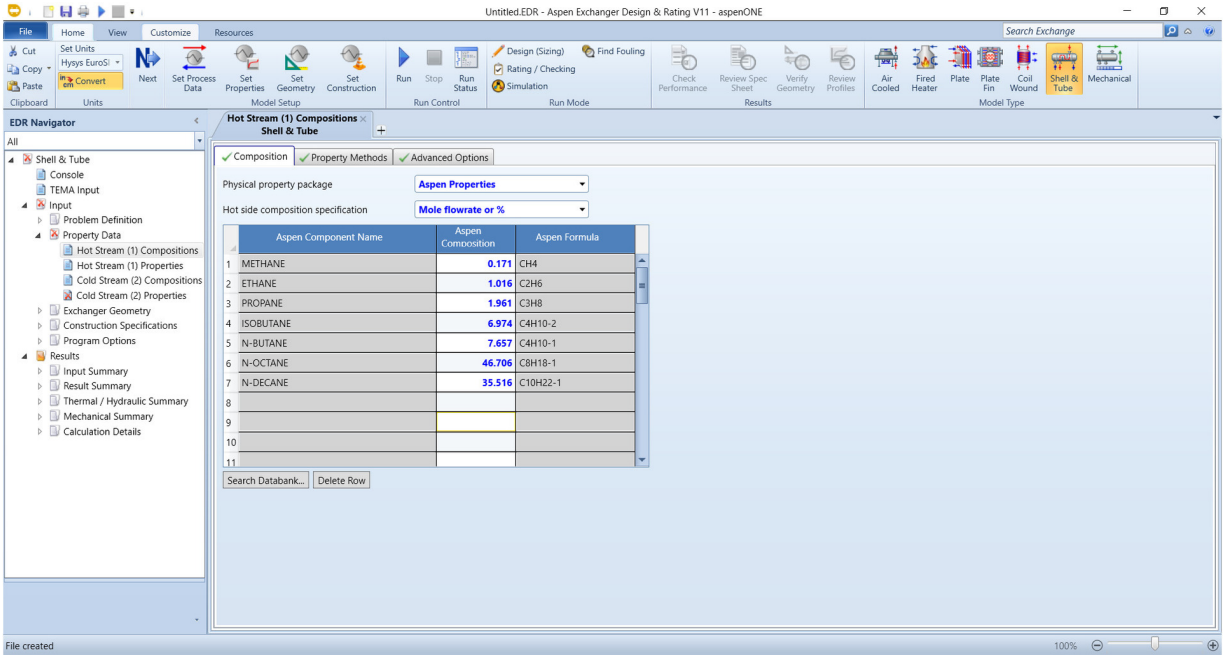


Figure 14.16: Heat exchanger sizing

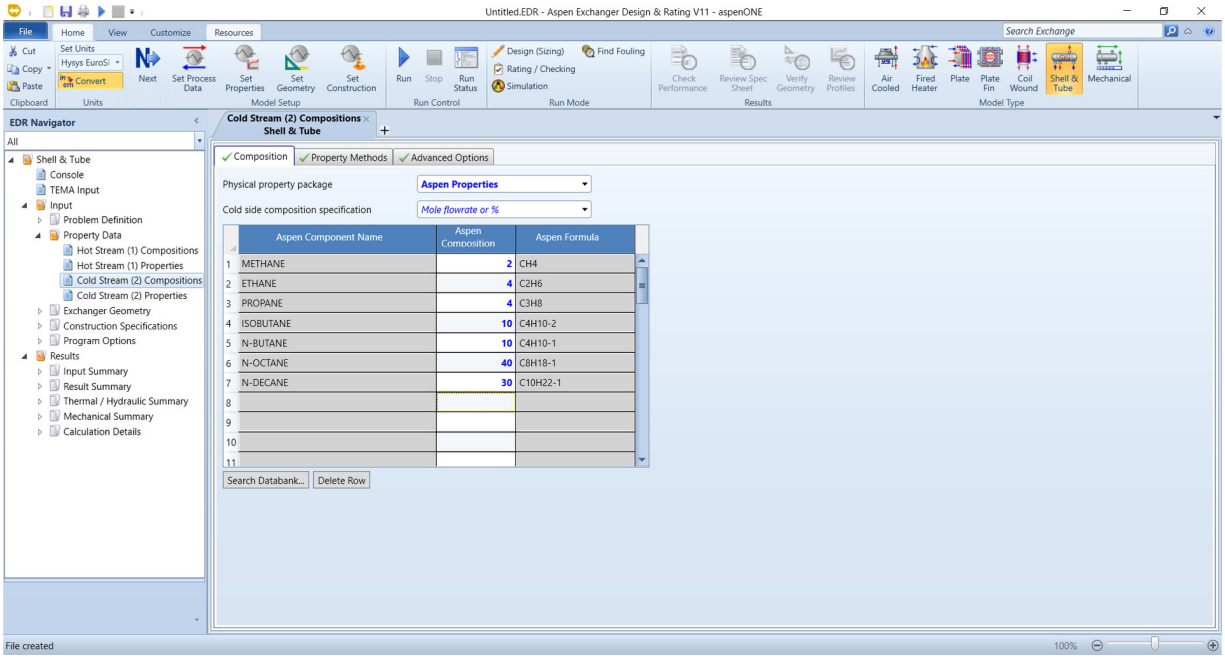


Figure 14.17: Heat exchanger sizing

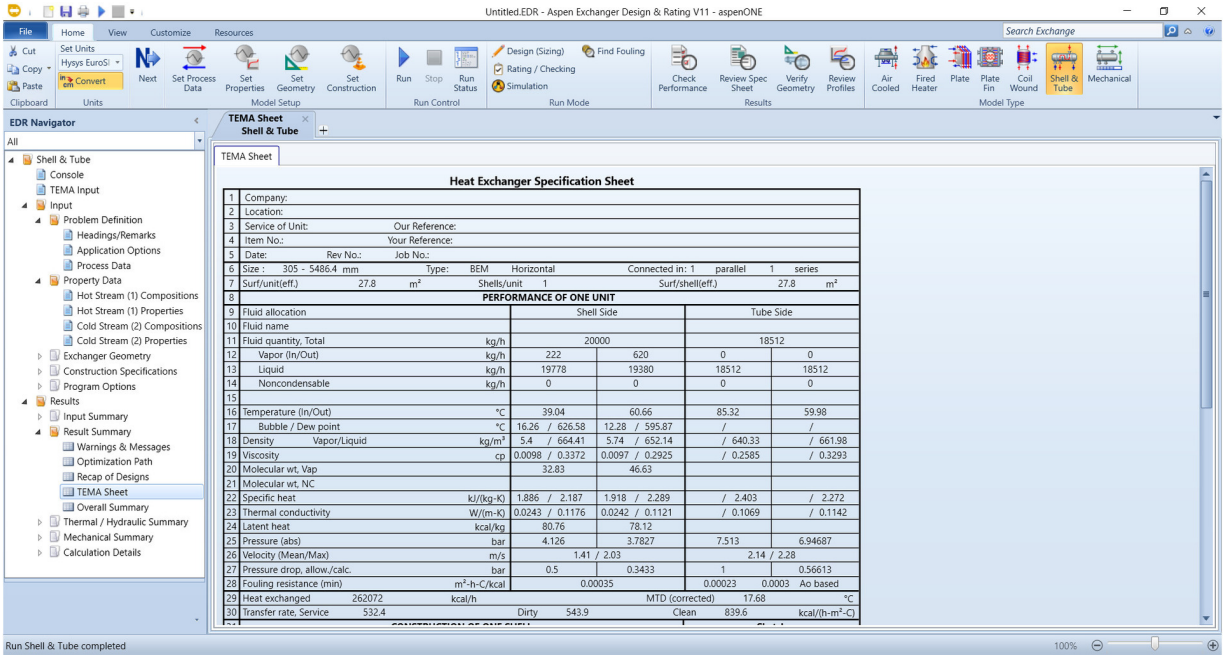
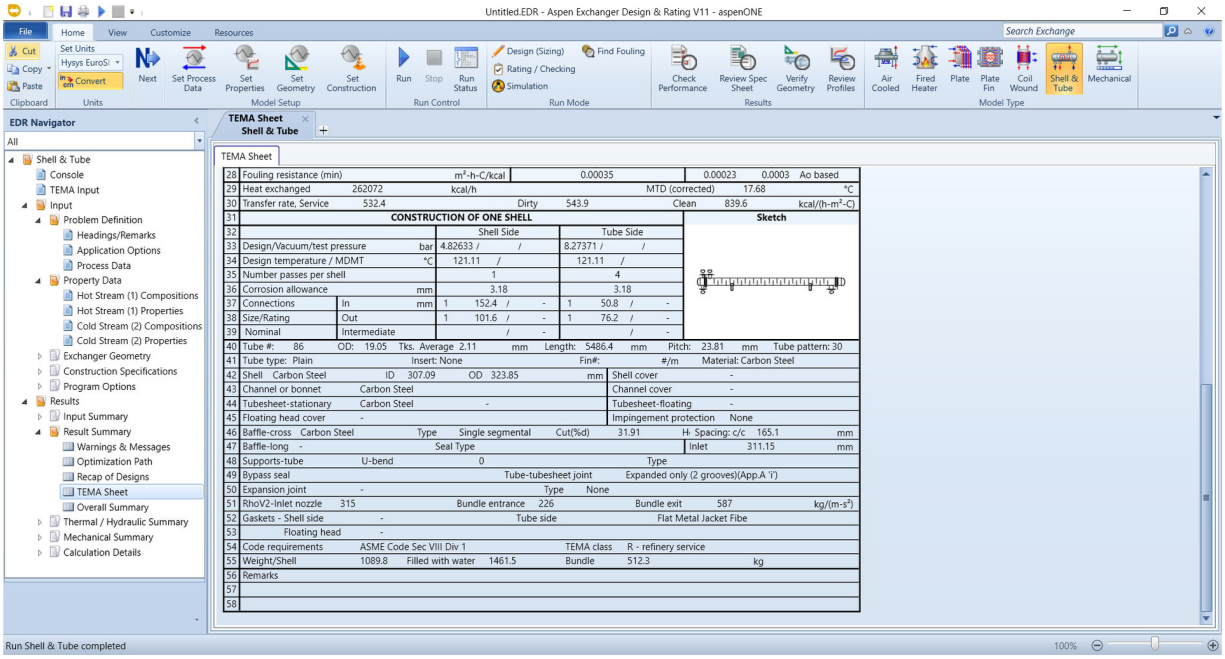


Figure 14.18: Heat exchanger sizing



Results

- HTA = 28 m^2
- No. of tubes = 86
- No. of passes = 4
- Tube length = 5000 mm
- Shell ID = 307 mm
- U_D = 622 W/m^2 .K

Pump sizing

Figure 14.19: Pump sizing

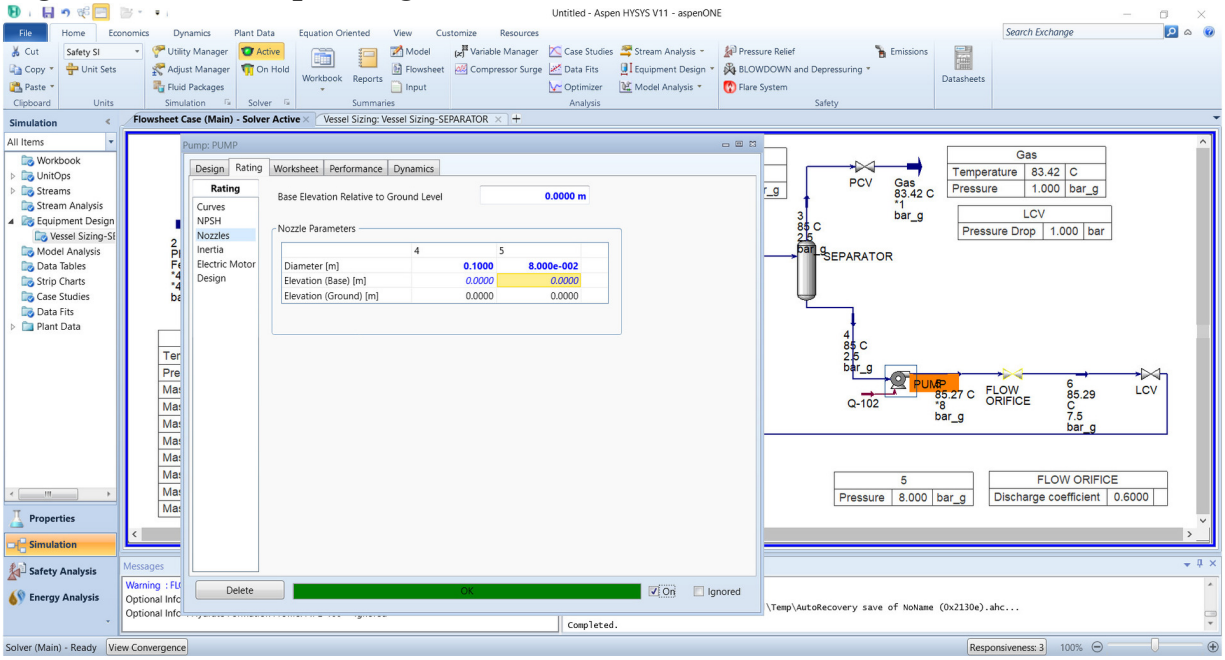


Figure 14.20: Pump sizing

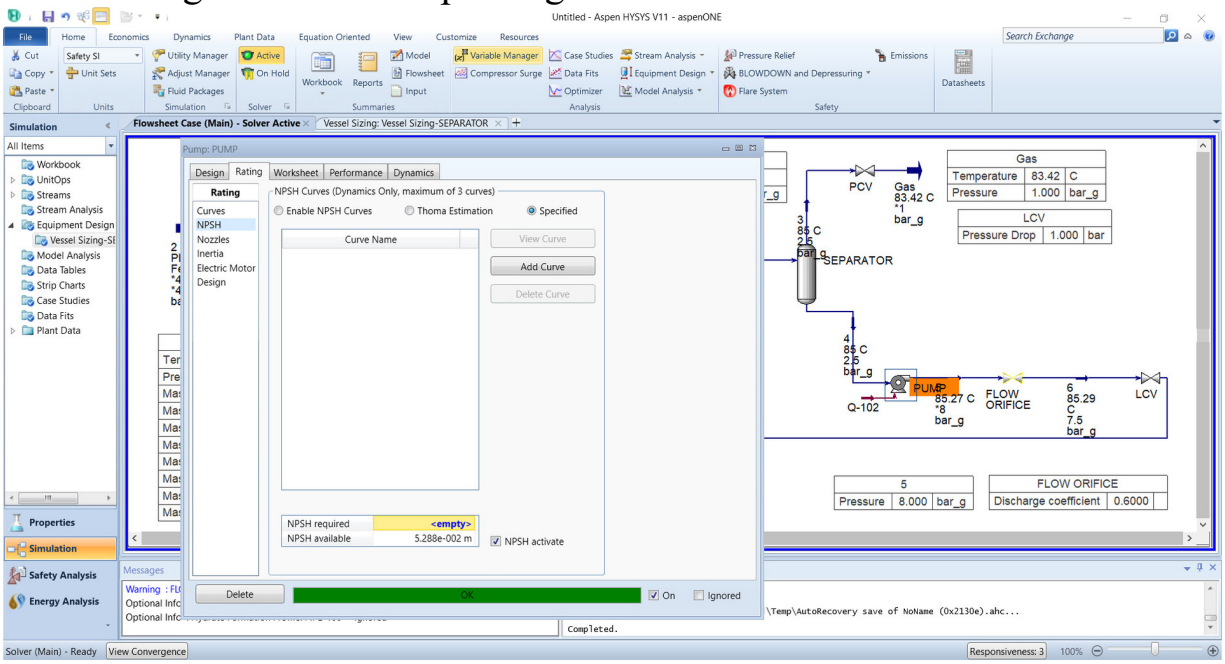
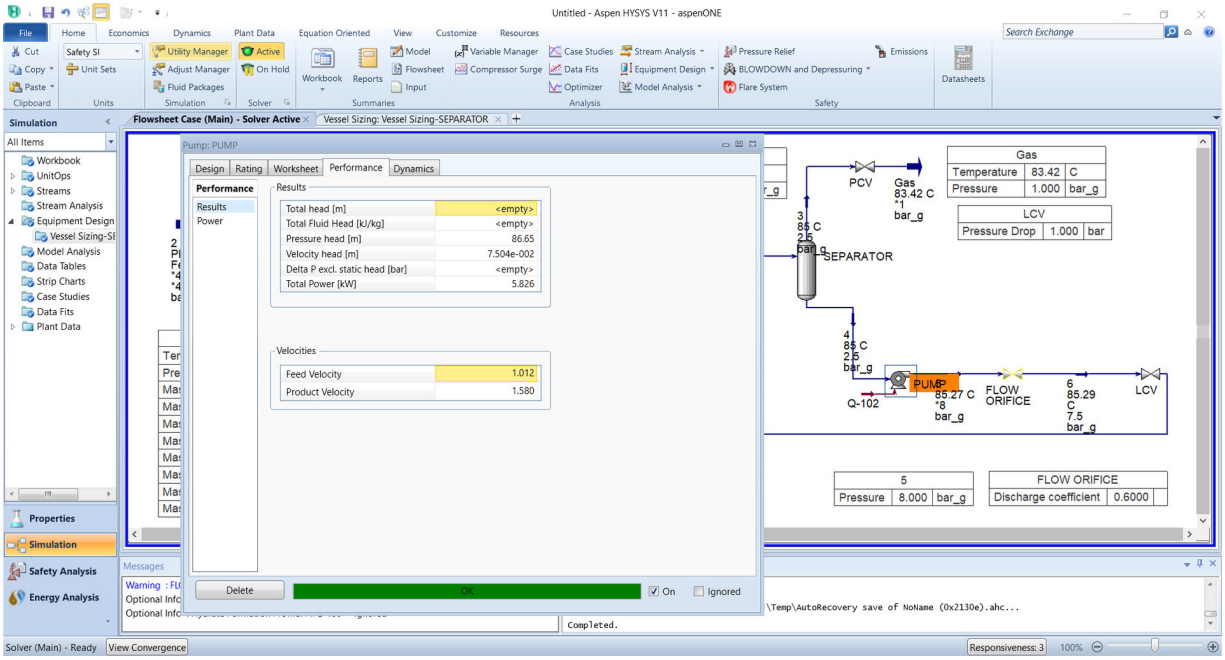


Figure 14.21: Pump sizing



Results

- Pump duty flow = 29 m^3 /h
- Pump head = 87 mlc

- $NPSH_A$ = 0.05 m
- Shaft power = 5.9 kW

Line sizing as per API 14E

- Two phase line size = 80 mm (v << Ve m/s)

Figure 14.22: Line sizing

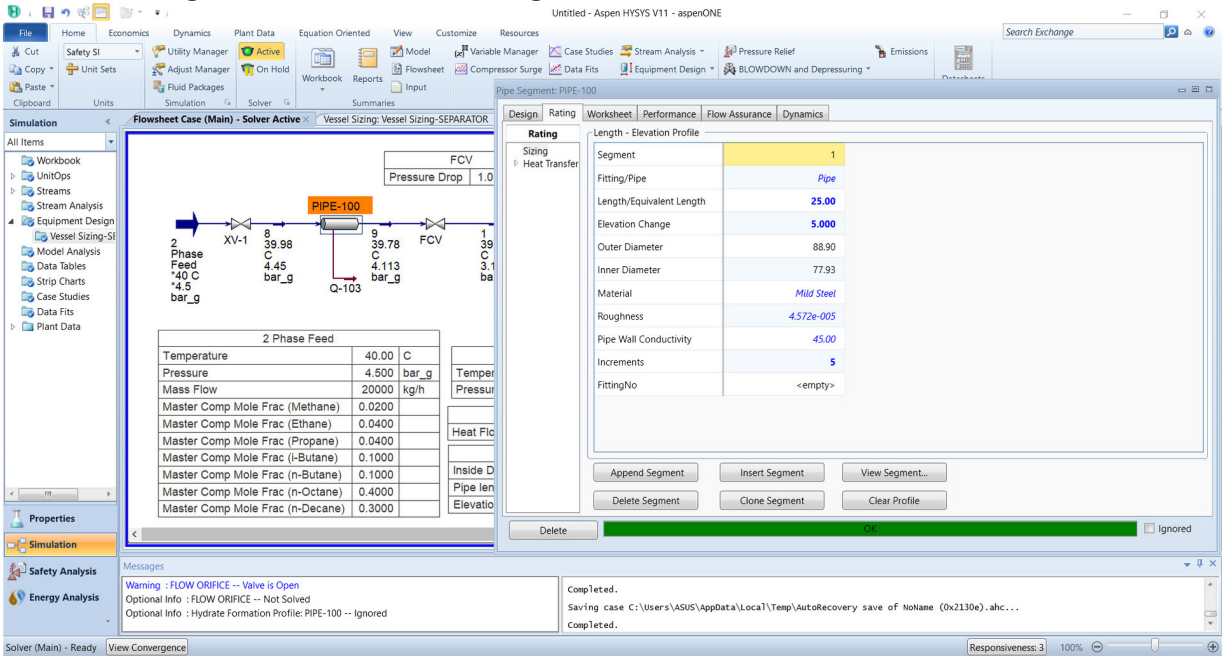
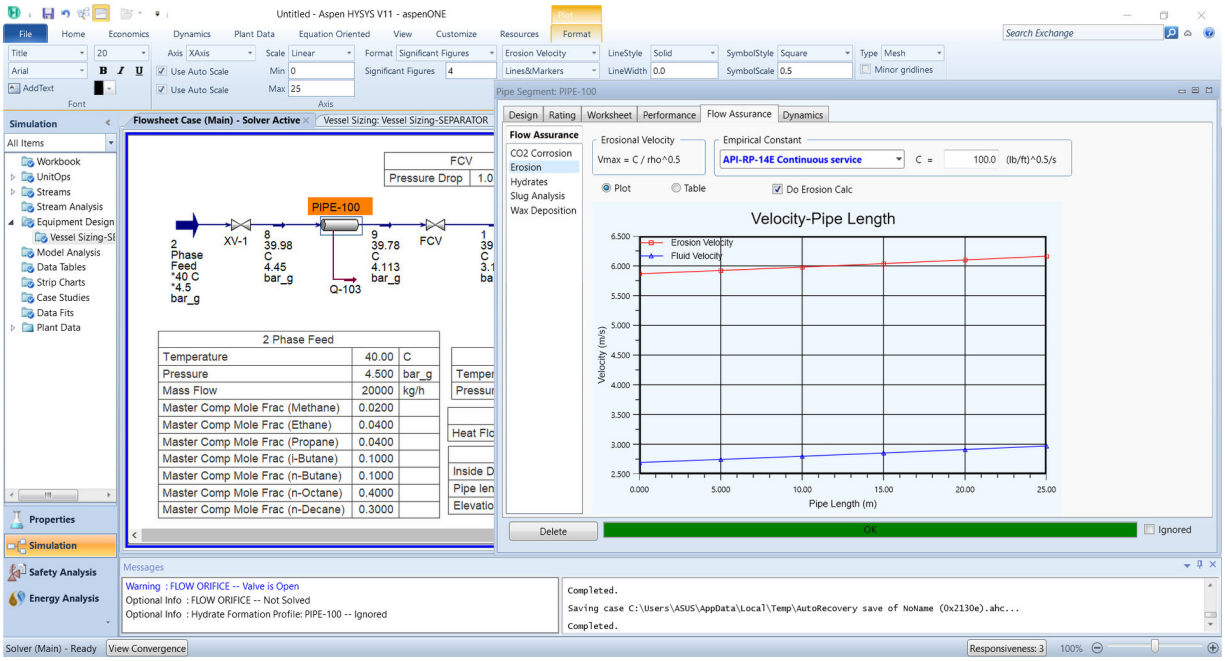


Figure 14.23: Line sizing



- Pump suction line size = 100 mm (v ~ 1 m/s)

Figure 14.24: Line sizing

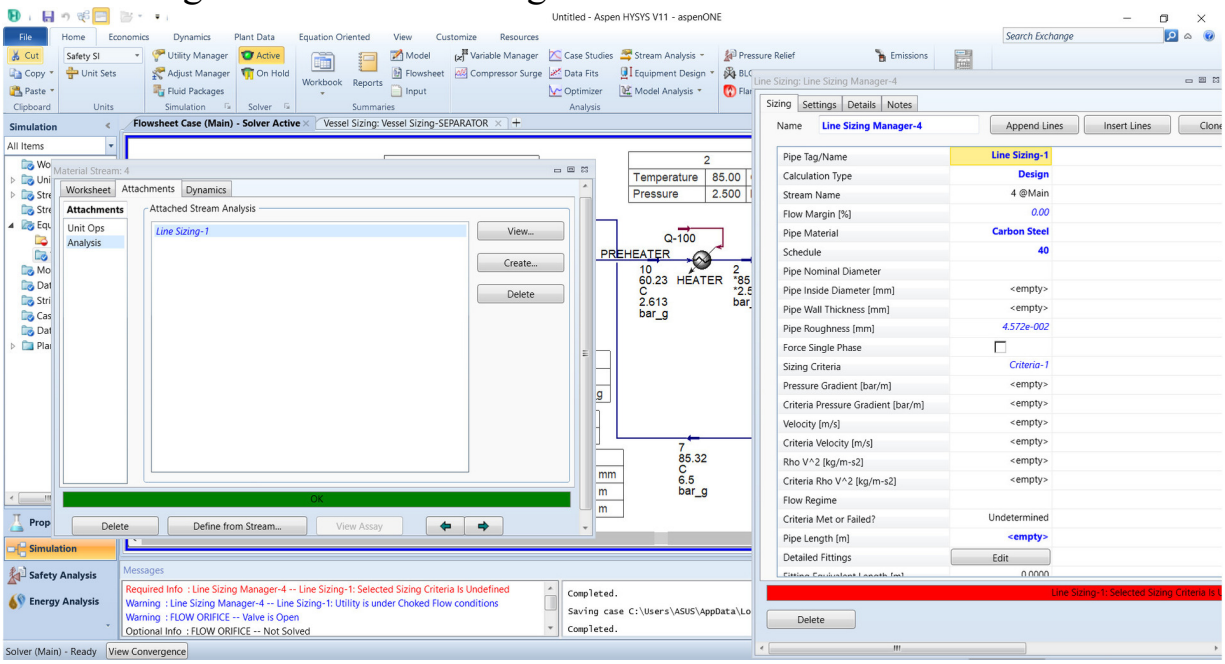
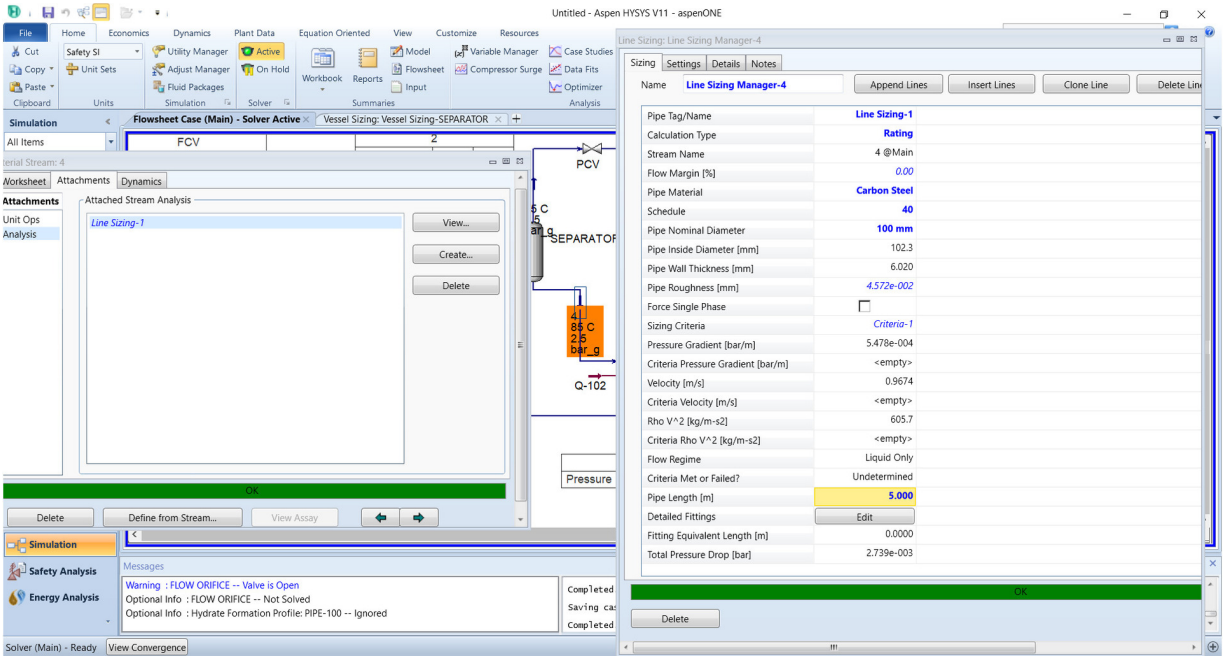


Figure 14.25: Line sizing



- Pump discharge line size = 80 mm (v ~ 2.7 m/s)

Figure 14.26: Line sizing

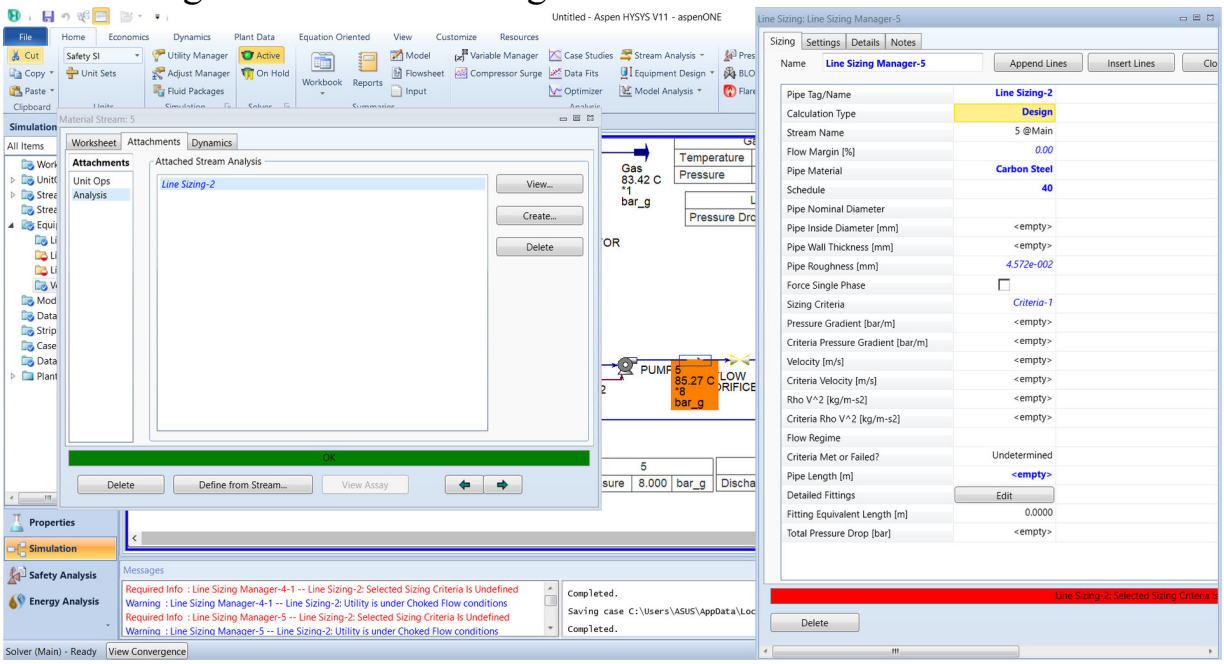
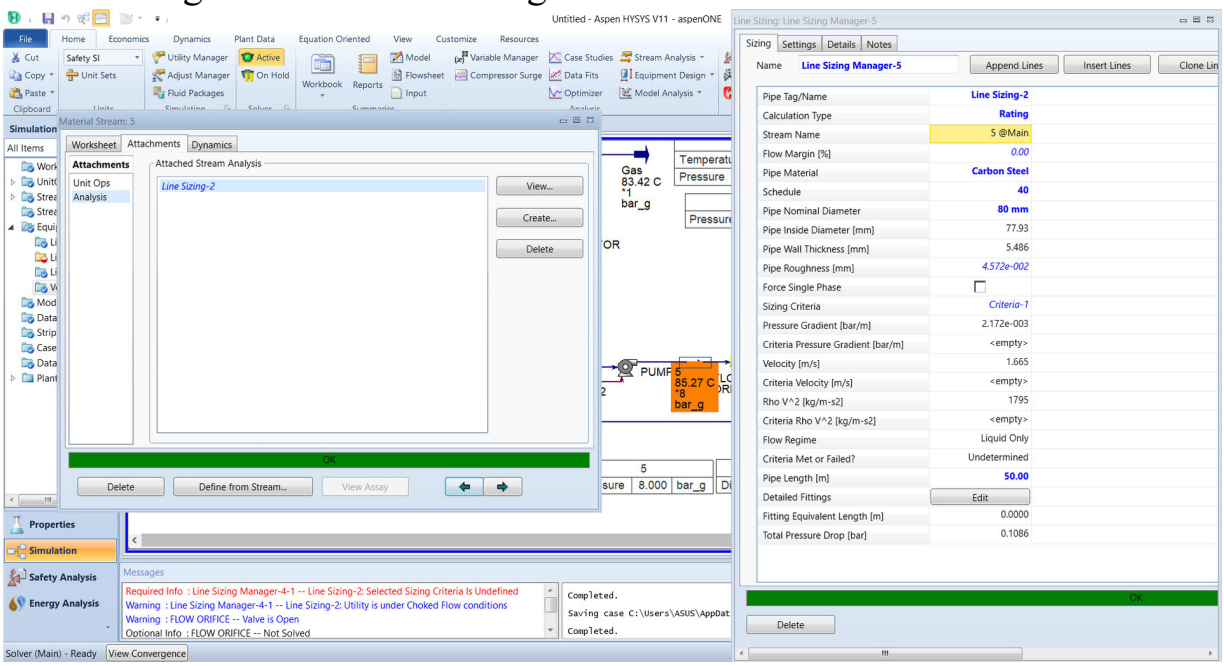


Figure 14.27: Line sizing



- Vapor line size = 100 mm (v < 18 m/s)

Figure 14.28: Line sizing

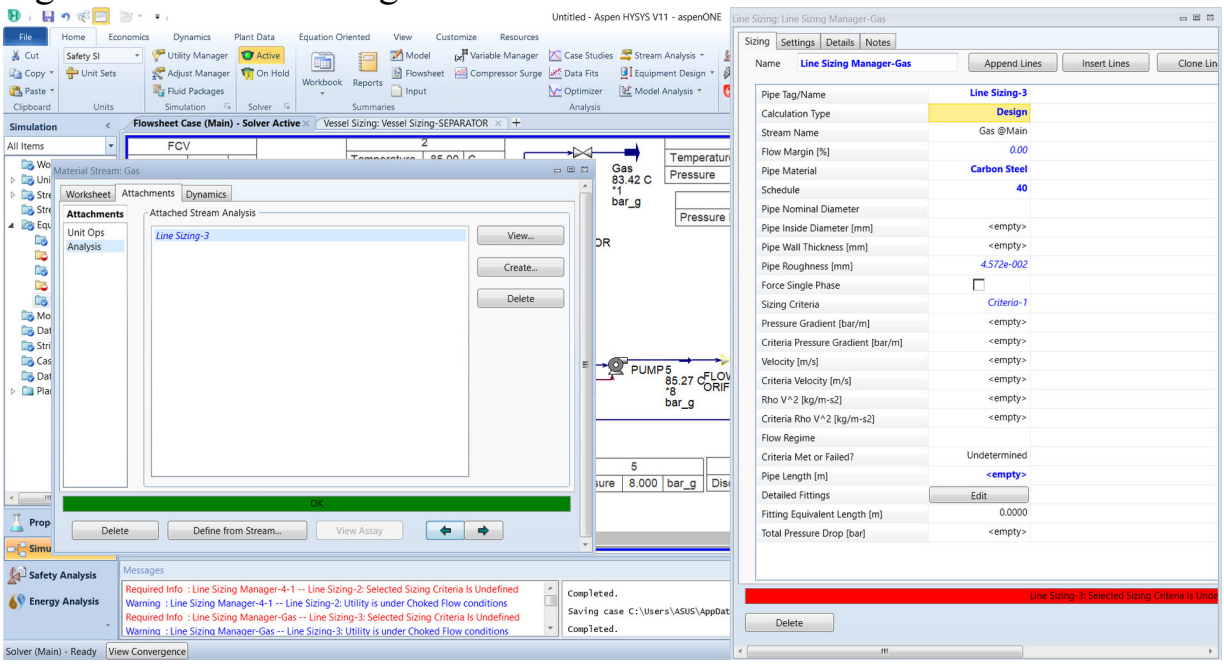
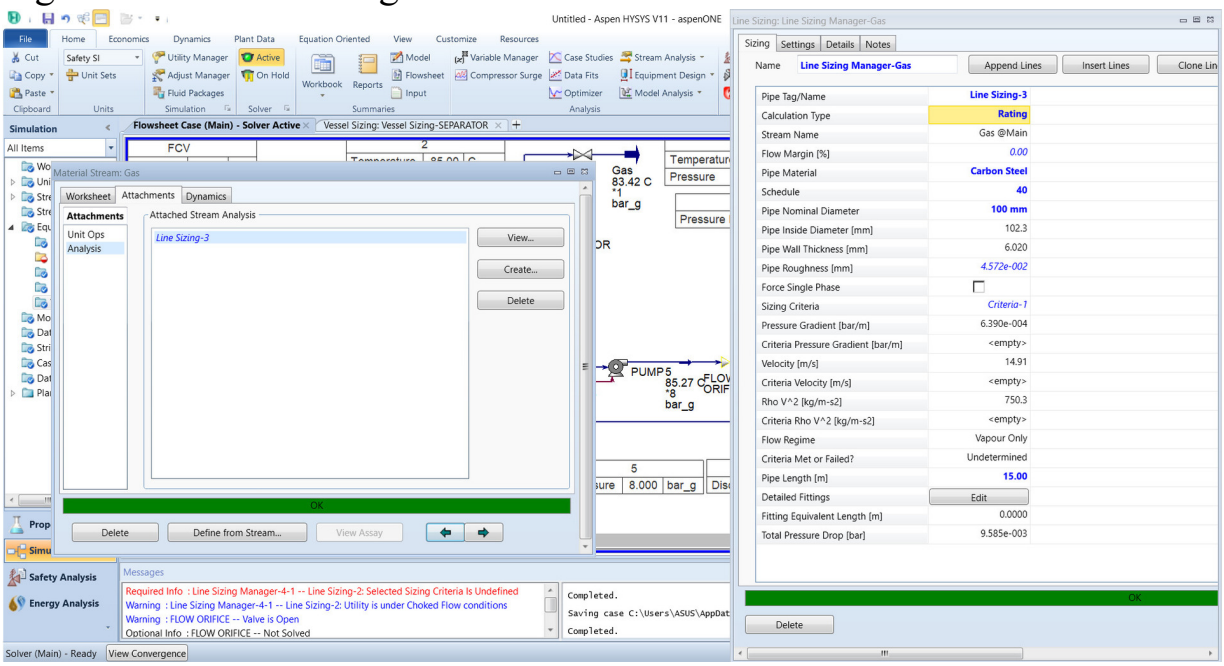


Figure 14.29: Line sizing



LCV sizing as per ISA 75.01.01

Figure 14.30: Control valve sizing

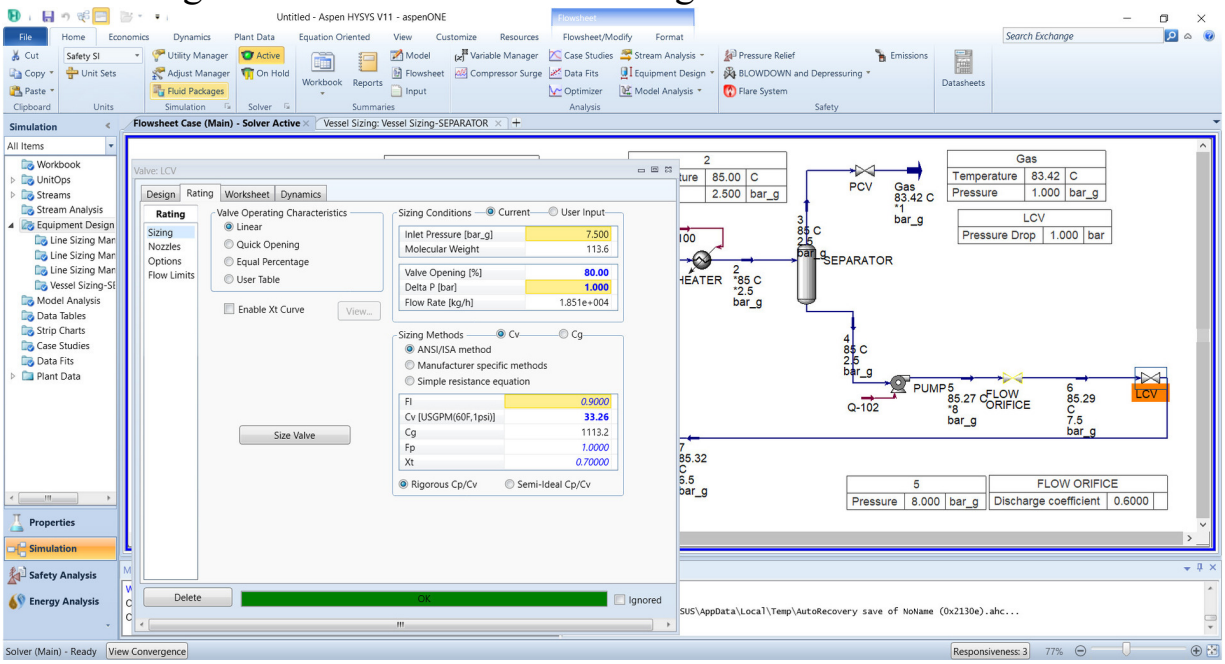
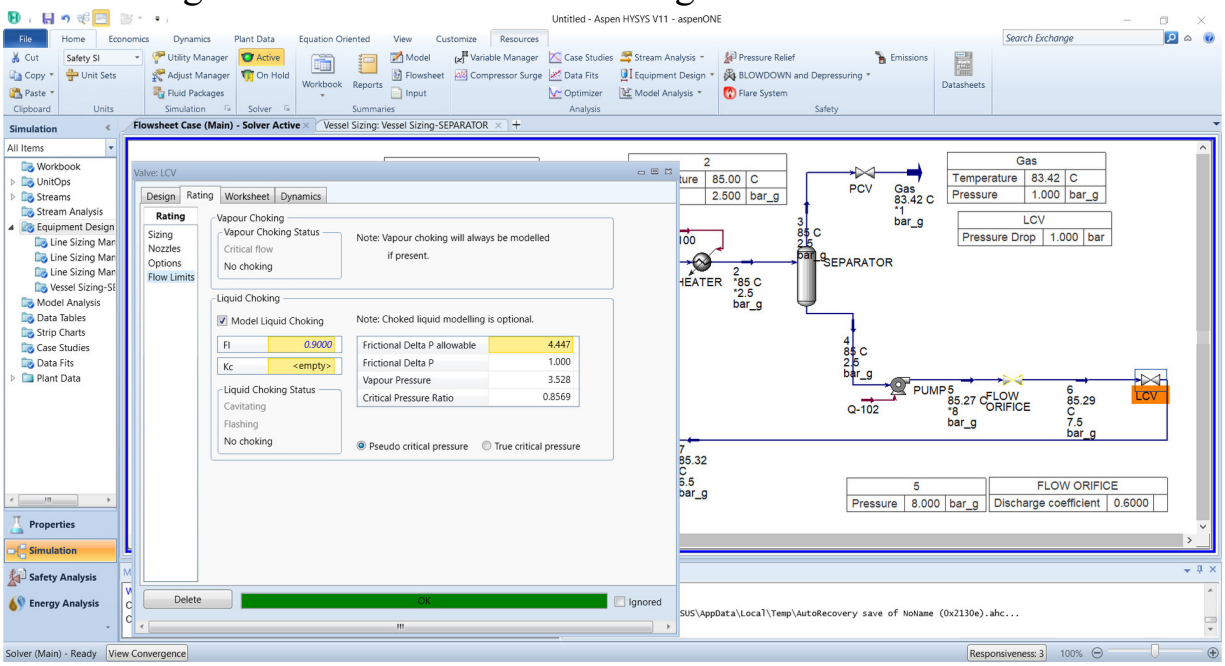


Figure 14.31: Control valve sizing

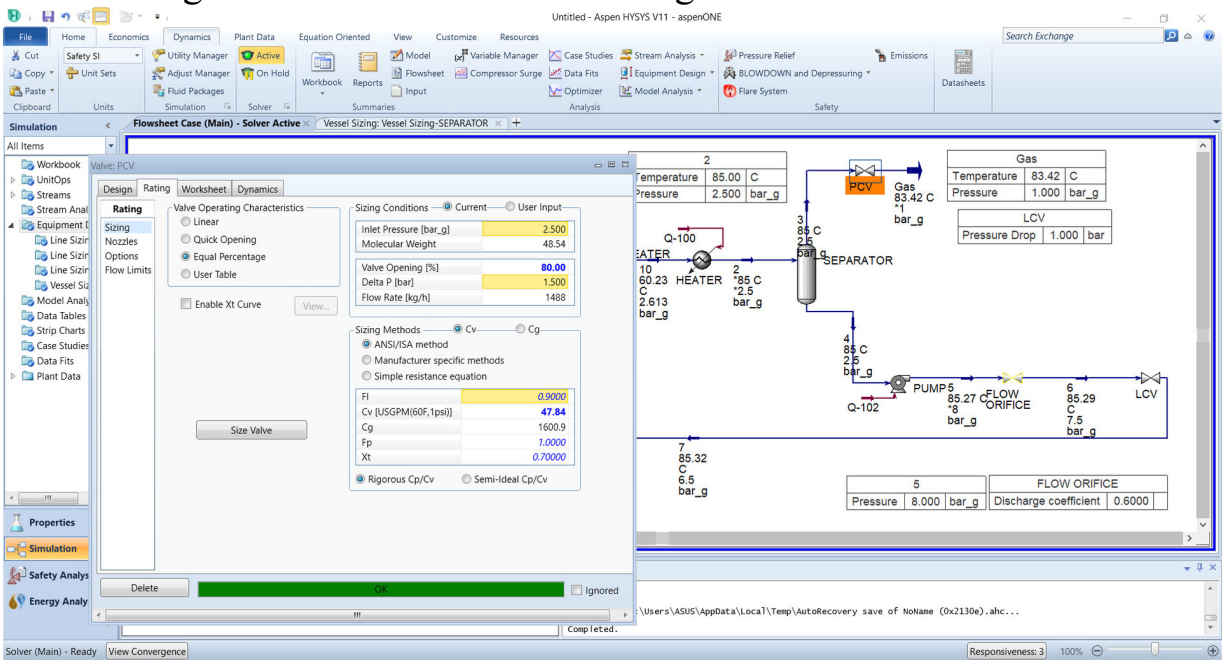


- Valve opening = 80%
- Pressure drop = 1 bar

- Cv = 34
- Valve size = 2” (Rule of thumb: Cv = 10 d^2)

PCV sizing as per ISA 75.01.01

Figure 14.32: Control valve sizing



- Valve opening = 80%
- Pressure drop = 1.5 bar
- Cv = 48
- Valve size = 2” (Rule of thumb: Cv = 10 d^2)

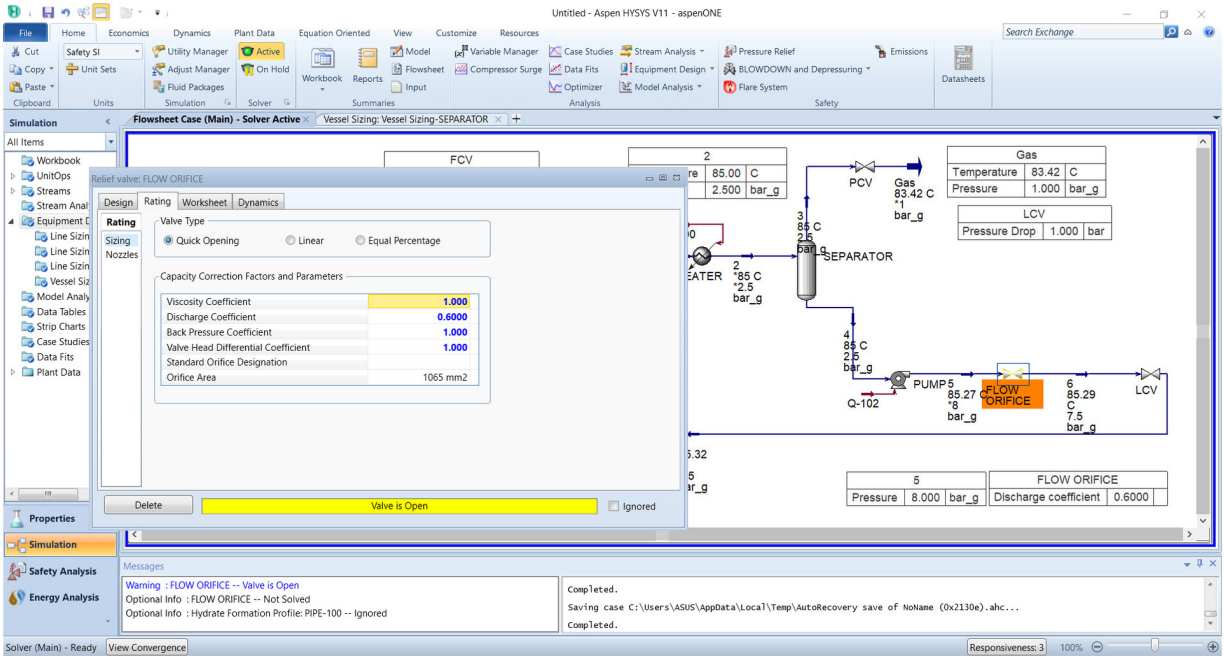
FCV sizing as per ISA 75.01.01

- Valve opening = 80%
- Pressure drop = 1 bar
- Cv = 58
- Valve size = 2.5” (Rule of thumb: Cv = 10 d^2)

Flow orifice sizing as per ISO 5167-2

Select the “Rating” tab and provide the “Discharge Coefficient” as 0.6.

Figure 14.33: Orifice sizing



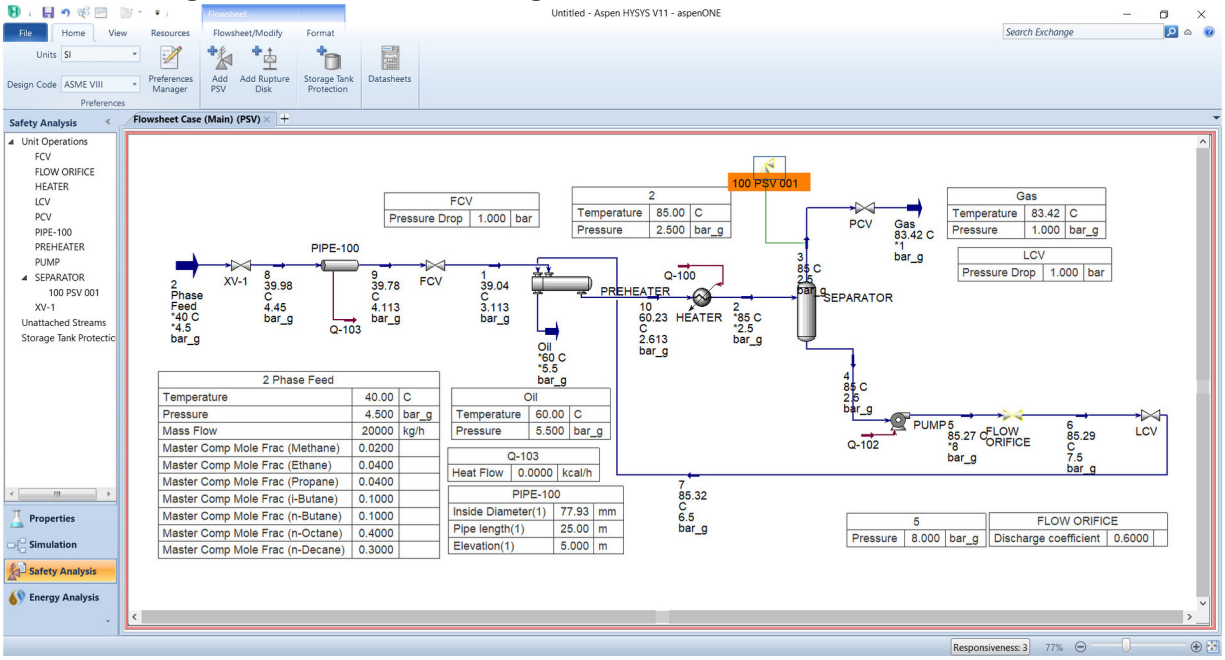
- Pressure drop = 0.5 bar
- Orifice area = 1065 mm^2

- Orifice diameter (d) = 37 mm
- Pipe ID (D) = 78 mm
- β ratio (d / D) = 0.47 (0.1 < β < 0.75)

PSV sizing as per API 520 Part 1 and API 526

Select “Safety Analysis” environment. “Add PSV” at the Separator top.

Figure 14.34: PSV sizing



Maximum operating pressure as per process is considered as 3.5 barg (normal operating pressure is 2.5 barg). Maximum operating pressure + 3.5 bar = 7 barg, this will be considered as process design pressure of separator. The same will be taken as PSV set pressure.

The normal operating temperature is 85 ° C. 30 ° C is added to 85 ° C and 120 ° C is considered as design temperature.

Figure 14.35: PSV sizing

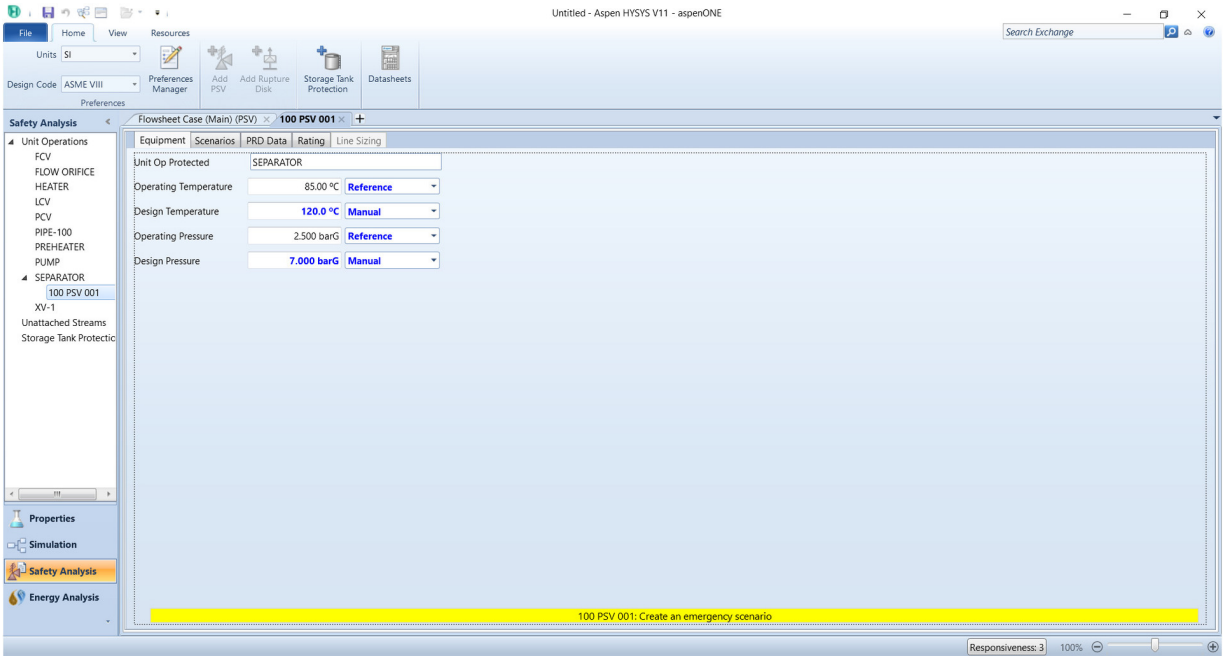


Figure 14.36: PSV sizing

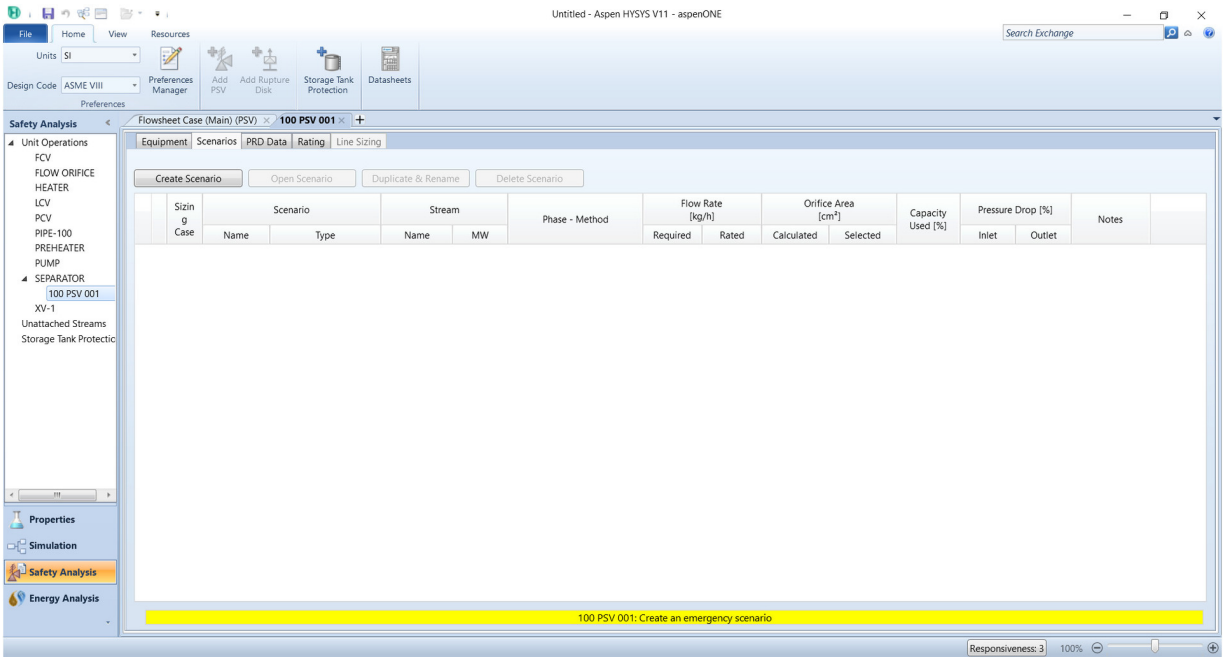

Untitled - Aspen HYSYS V11 - aspenONE
File
Home
View
Resources
Search Exchange
Units
SI
Design Code
ASME VIII
Preferences
Preferences Manager
Add PSV
Add Rupture Disk
Storage Tank Protection
Datasheets
Safety Analysis
Unit Operations
FCV
FLOW ORIFICE
HEATER
LCV
PCV
PIPE-100
PREHEATER
PUMP
SEPARATOR
100 PSV 001
XV-1
Unattached Streams
Storage Tank Protectic
Properties
Simulation
Safety Analysis
Energy Analysis
Flowsheet Case (Main) (PSV)
100 PSV 001
Equipment
Scenarios
PRD Data
Rating
Line Sizing
Create Scenario
Open Scenario
Duplicate & Rename
Delete Scenario
Sizing Case
Scenario
Name
Type
Stream
Name
MW
Phase - Method
Flow Rate [kg/h]
Required
Rated
Orifice Area [cm²]
Calculated
Selected
Capacity Used [%]
Pressure Drop [%]
Inlet
Outlet
Notes
100 PSV 001: Create an emergency scenario
Responsiveness: 3
100%

Figure 14.37: PSV sizing

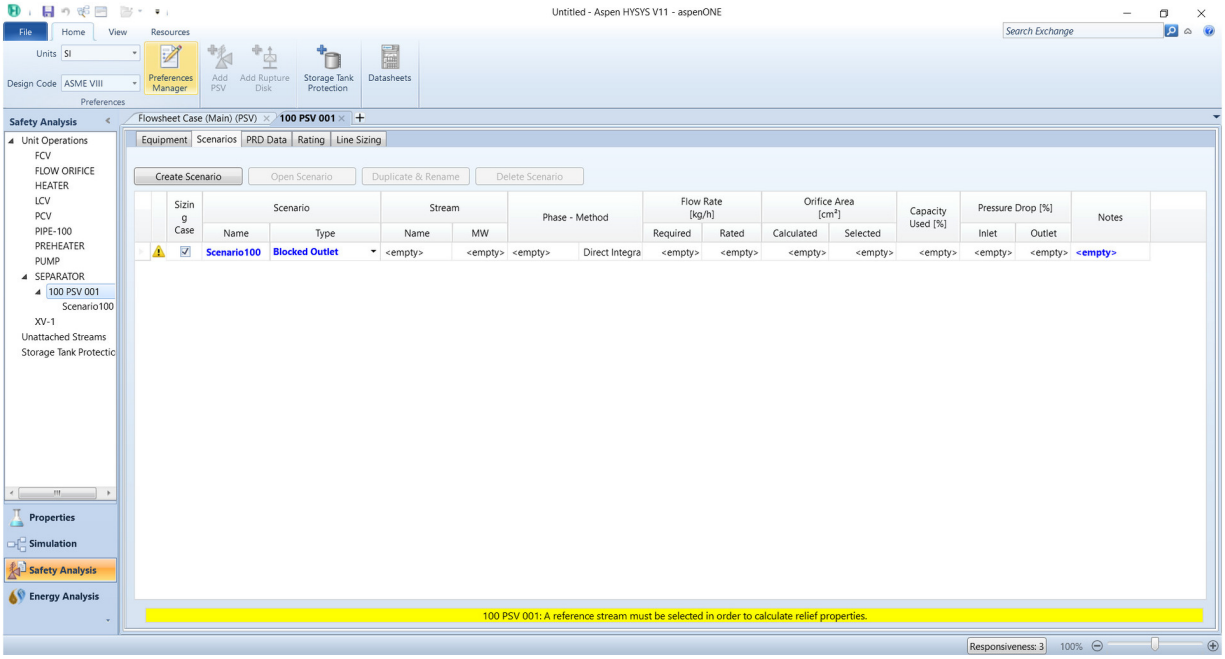


Figure 14.38: PSV sizing

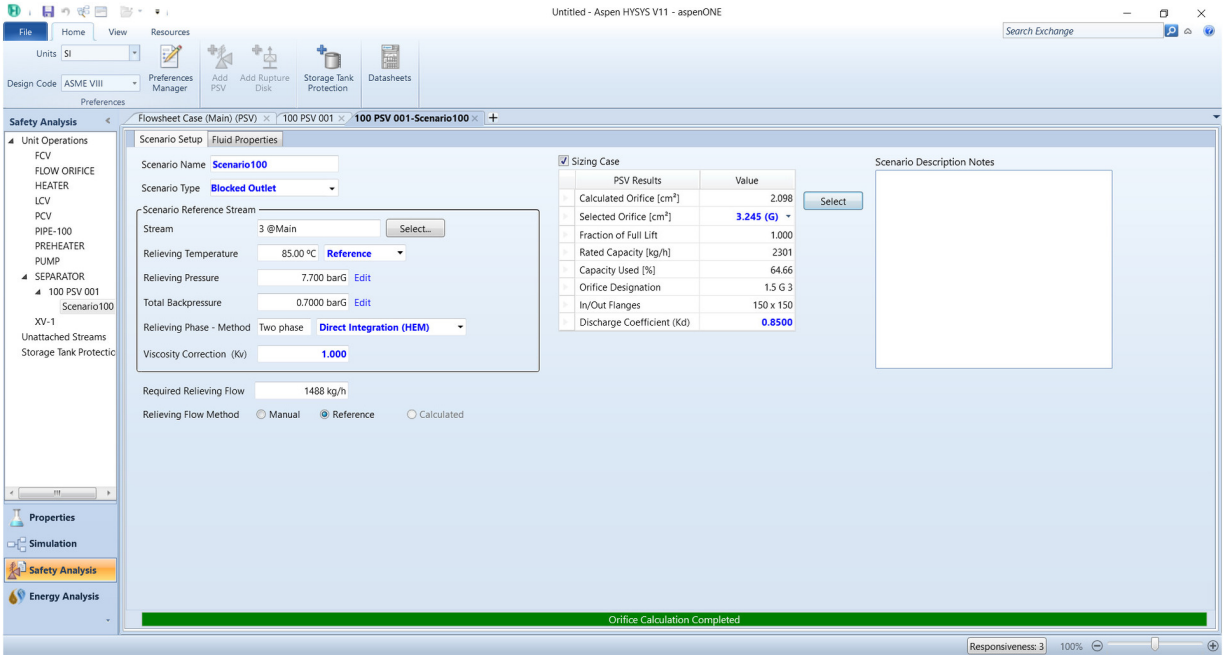
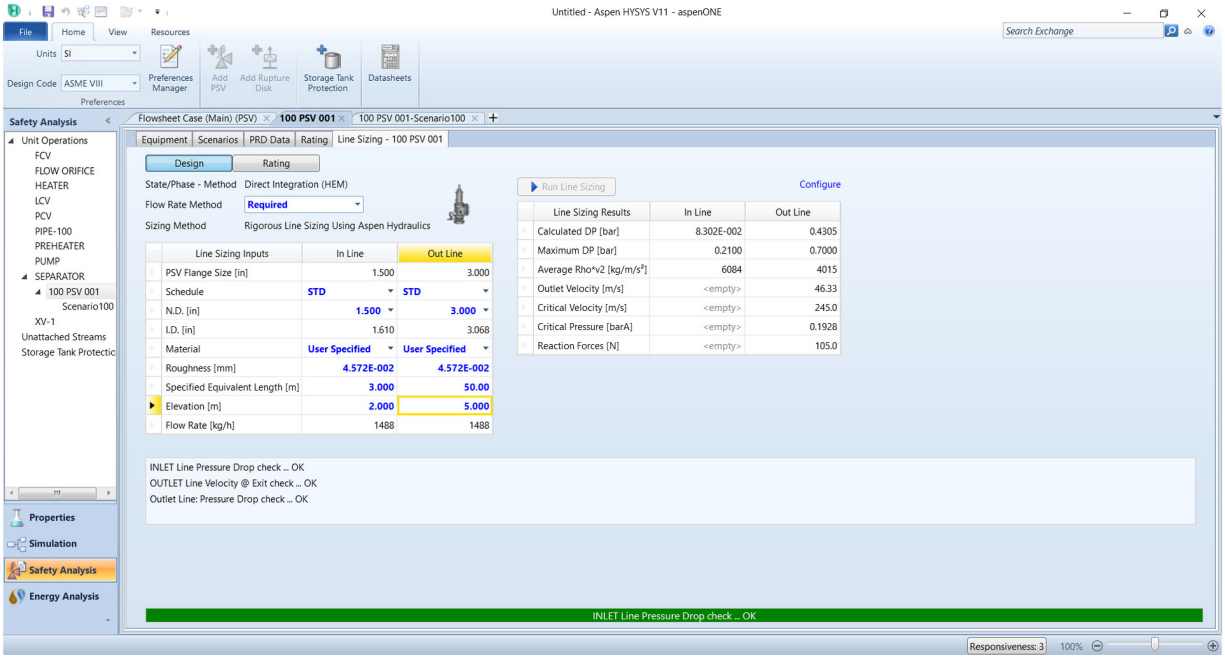


Figure 14.39: PSV sizing



- Set pressure = 6.5 barg
- PSV size = 1.5G3
- Inlet line size = 1.5”
- Outlet line size = 3”

Exercises

a. Mr. Vishal needs to design a separation system to separate gas from oil in a 2 phase fluid, that is available at 40 ° C, 4.5 barg pressure, and 30000 kg/h. Mole fraction of feed is provided below;

- Methane : 0.04
- Ethane : 0.02
- Propane : 0.04
- i-Butane : 0.15
- n-Butane : 0.05
- n-Octane : 0.40
- n-Decane : 0.30

The gaseous product shall be connected to the existing LP flare header at 1.5 barg pressure. Oil shall be sent to the existing header at 50 ° C and 4.5 barg pressure. RVP of the oil shall be min. 0.045 barg.
Please help Mr. Vishal to carry out the plant design. Also, prepare BOQ and provide an approximate cost estimate as well.

b. Mr. Rana needs to design a separation system to separate gas from oil in a 2 phase fluid, that is available at 40 ^{o} C, 4.5 barg pressure, and 90000 kg/h. Mole fraction of feed is provided below;

- Methane : 0.04
- Ethane : 0.02
- Propane : 0.04
- i-Butane : 0.15
- n-Butane : 0.05
- n-Octane : 0.40
- n-Decane : 0.30

The gaseous product shall be connected to the existing LP flare header at 1.5 barg pressure. Oil shall be sent to the existing header at 25 ^{o} C and 4.5 barg pressure. RVP of the oil shall be min. 0.045 barg.

Please help Mr. Rana to carry out the plant design. Also, prepare BOQ and provide an approximate cost estimate as well.

CHAPTER 15: ASPEN HYSYS DYNAMICS (SIMPLE PROBLEM)

In this chapter, we are going to discuss the dynamic modeling of two-phase separator, which includes modeling of the material stream, separator, valves, and PID block. Also, one will be able to manipulate the variables and check the dynamic response.

Once you have completed this module, you will be able to:

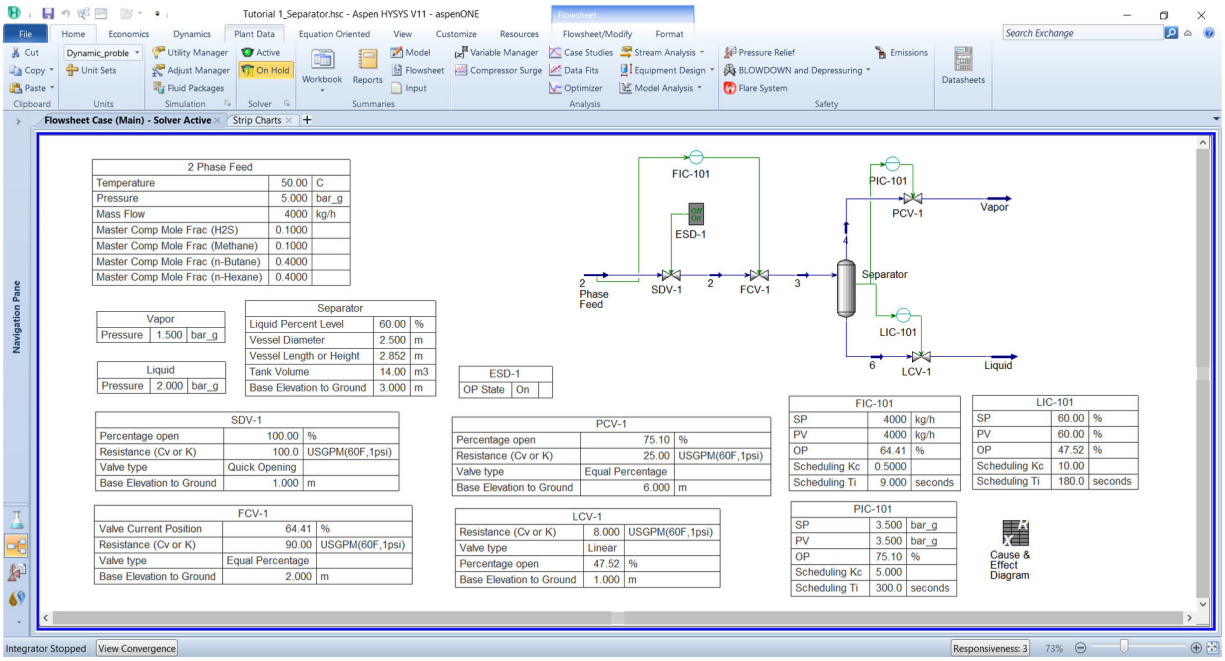
- Simulate valve in dynamic mode.
- Simulate vessel in dynamic mode.
- Simulate controller in dynamic mode.
- Study the dynamic behavior of the model.
- Prepare Cause and Effect Matrix

Please refer HYSYS model snapshot below;

A two-phase feed stream is sent to the separator.

- SDV-1 acts as an emergency shutdown valve. A negligible pressure drop of 5 kPa is to be considered for the same.
- FCV- controls the mass flow rate of the feed stream. The pressure drop of 150 kPa is to be considered for the same.
- Separator separates vapor and liquid streams.
- PCV- controls the separator pressure.
- LCV- controls the separator level.
- Vapor stream tie-in point pressure is 1.5 barg and that of liquid stream is 2 barg.
- Use Peng Robinson as a fluid package.

Figure 15.1: Steady-state model



2 Phase Feed		
Temperature	50.00	C
Pressure	5.000	bar_g
Mass Flow	4000	kg/h
Master Comp Mole Frac (H2S)	0.1000	
Master Comp Mole Frac (Methane)	0.1000	
Master Comp Mole Frac (n-Butane)	0.4000	
Master Comp Mole Frac (n-Hexane)	0.4000	

Vapor		
Pressure	1.500	bar_g

Liquid		
Pressure	2.000	bar_g

Separator		
Liquid Percent Level	60.00	%
Vessel Diameter	2.500	m
Vessel Length or Height	2.852	m
Tank Volume	14.00	m3
Base Elevation to Ground	3.000	m

ESD-1		
OP State	On	

SDV-1		
Percentage open	100.00	%
Resistance (Cv or K)	100.0	USGPM(60F,1psi)
Valve type	Quick Opening	
Base Elevation to Ground	1.000	m

PCV-1		
Percentage open	75.10	%
Resistance (Cv or K)	25.00	USGPM(60F,1psi)
Valve type	Equal Percentage	
Base Elevation to Ground	6.000	m

FIC-101		
SP	4000	kg/h
PV	4000	kg/h
OP	64.41	%
Scheduling Kc	0.5000	
Scheduling Ti	9.000	seconds

LIC-101		
SP	60.00	%
PV	60.00	%
OP	47.52	%
Scheduling Kc	10.00	
Scheduling Ti	180.0	seconds

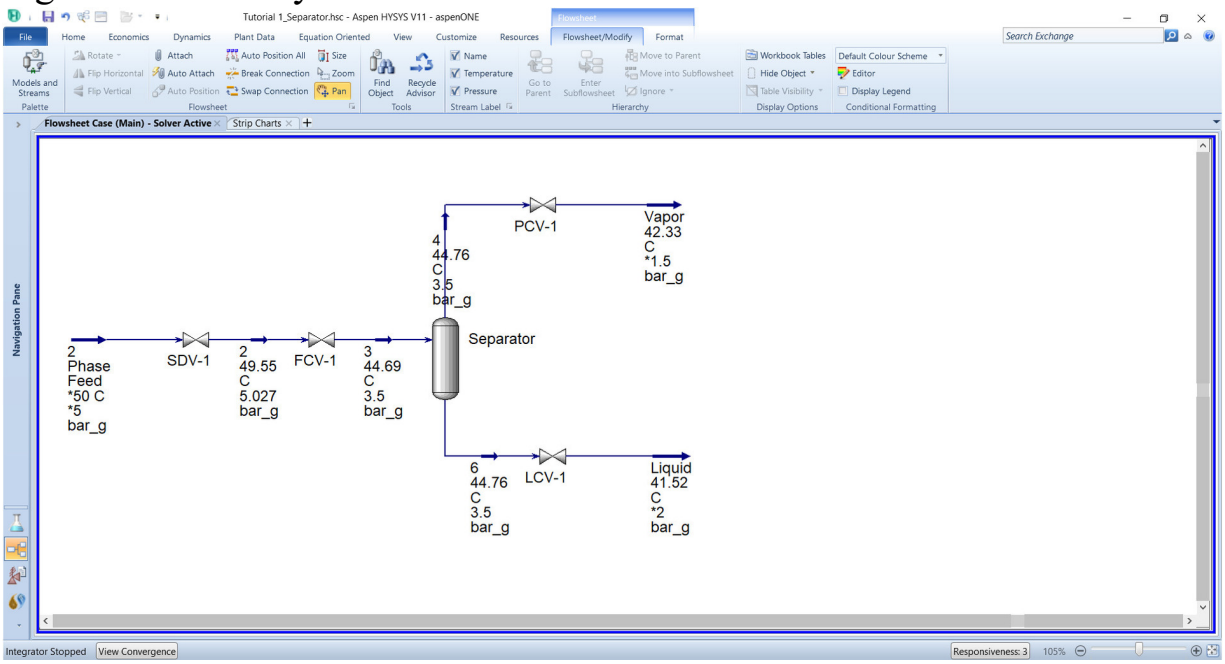
FCV-1		
Valve Current Position	64.41	%
Resistance (Cv or K)	90.00	USGPM(60F,1psi)
Valve type	Equal Percentage	
Base Elevation to Ground	2.000	m

LCV-1		
Resistance (Cv or K)	8.000	USGPM(60F,1psi)
Valve type	Linear	
Percentage open	47.52	%
Base Elevation to Ground	1.000	m

PIC-101		
SP	3.500	bar_g
PV	3.500	bar_g
OP	75.10	%
Scheduling Kc	5.000	
Scheduling Ti	300.0	seconds

Add components and fluid package.
Prepare the model as shown below.

Figure 15.2: Steady-state model



Select dynamic specifications as shown below; Flow rate is calculated w.r.to pressure differential.

Figure 15.3: Dynamic specifications

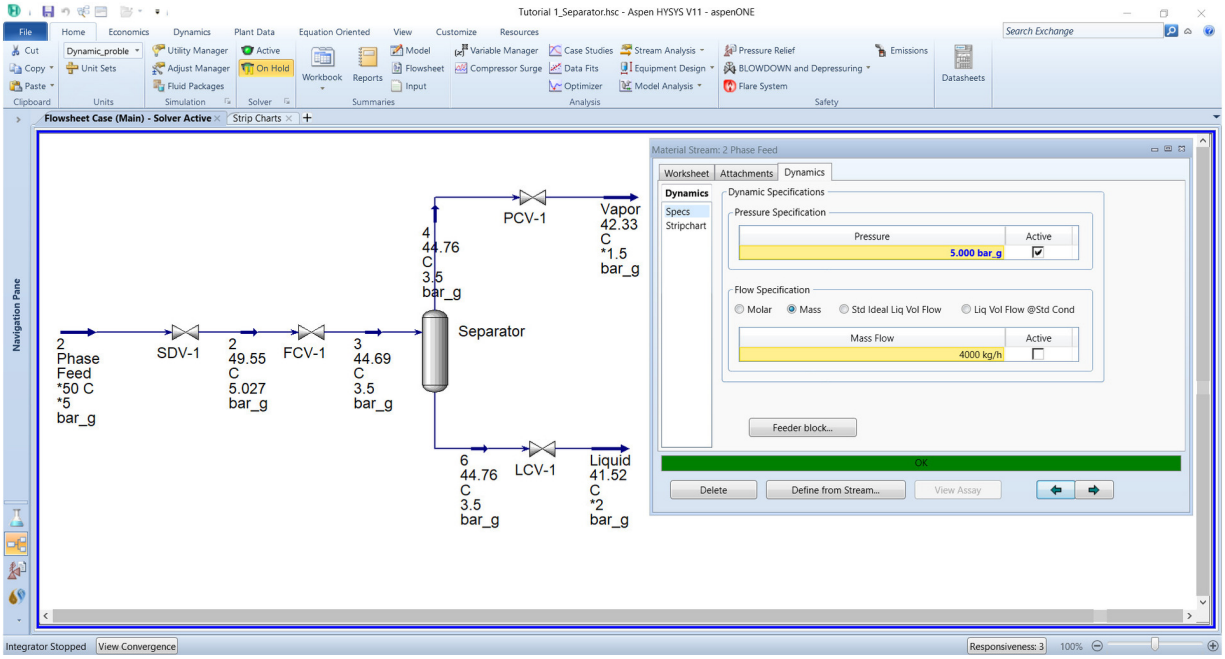


Figure 15.4: Dynamic specifications

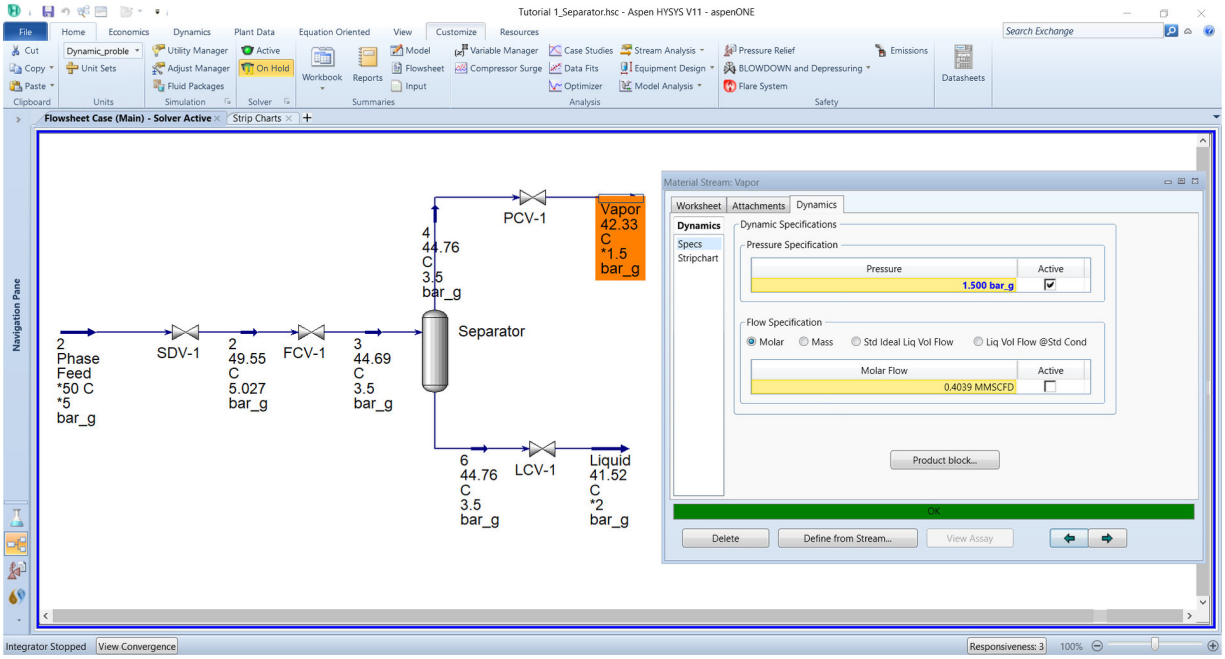
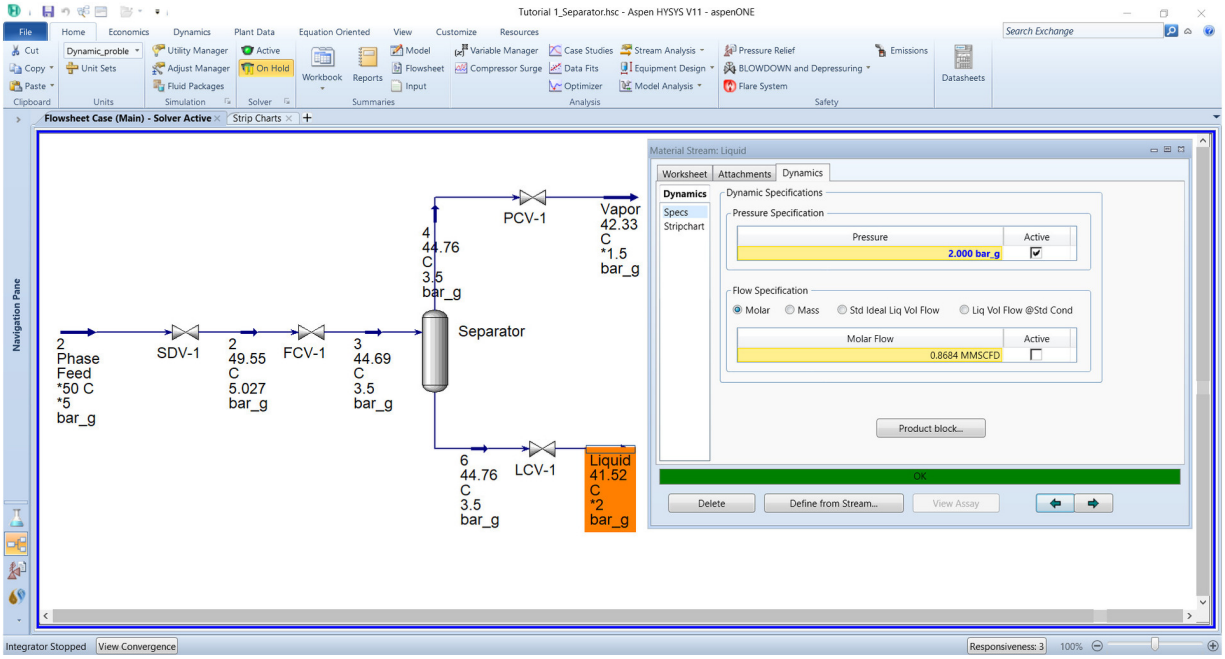


Figure 15.5: Dynamic specifications



Provide FIC-1, PIC-1, and LIC-1.

Figure 15.6: PID controller

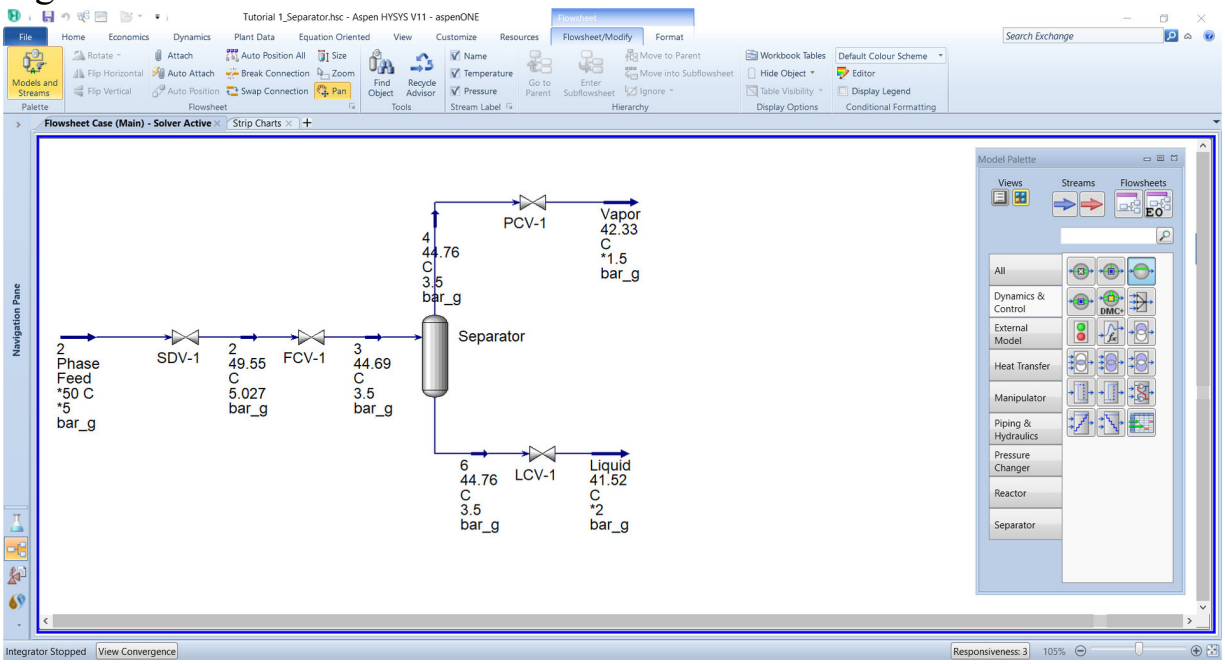
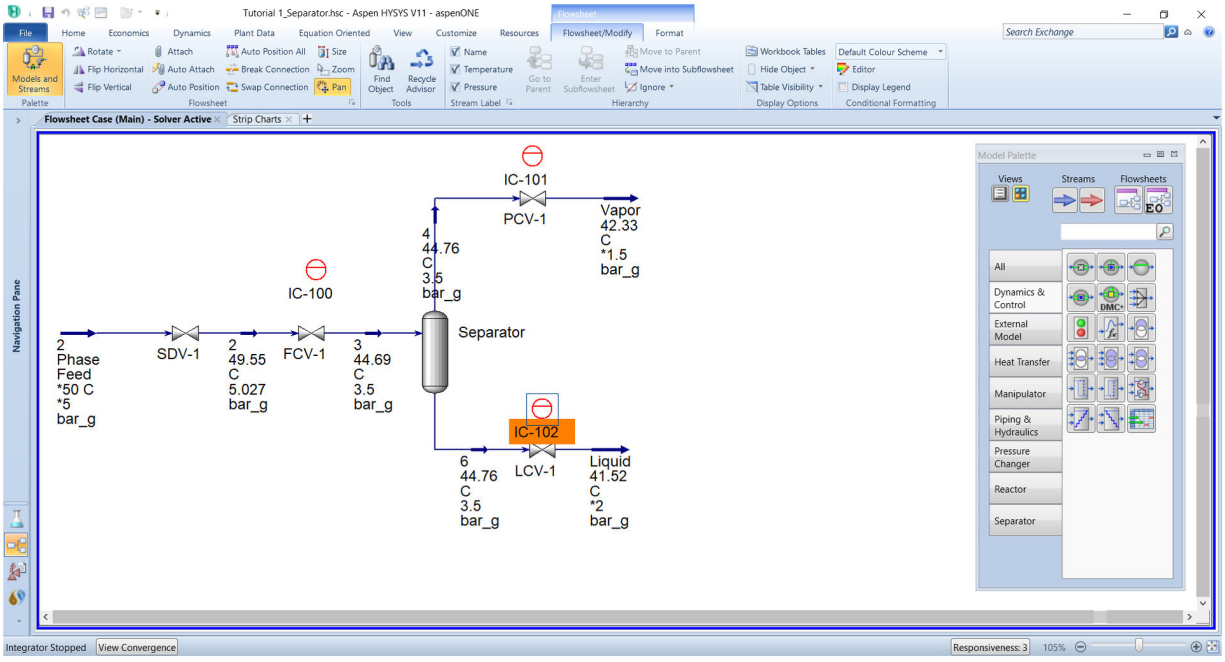
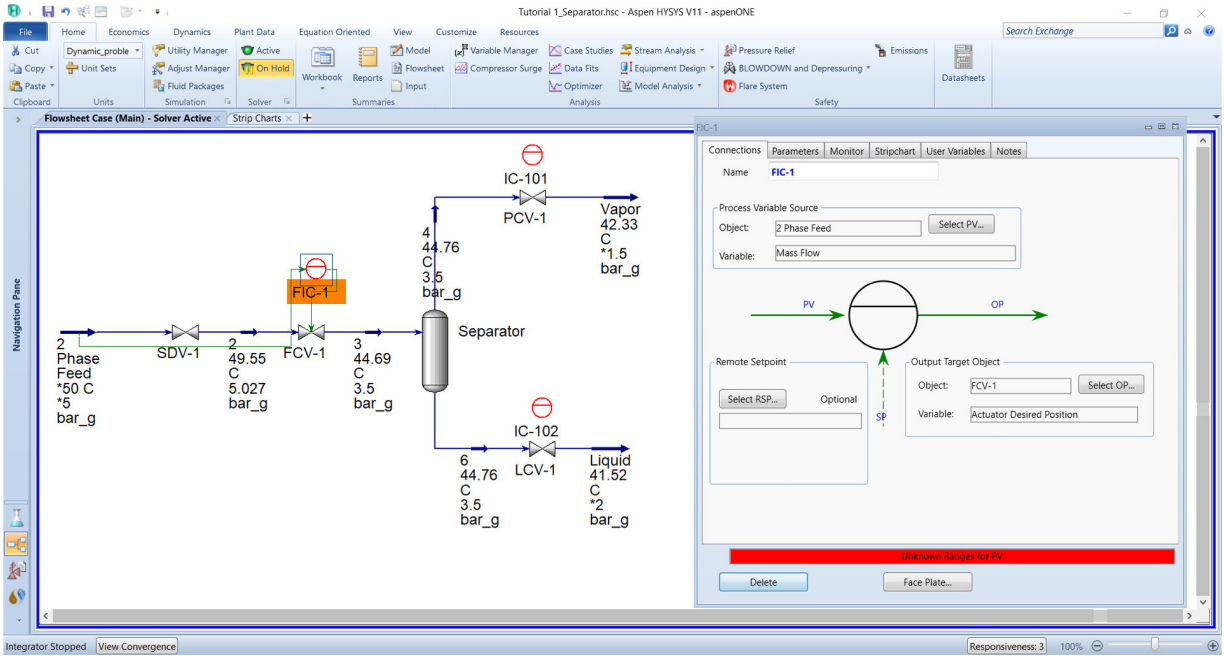


Figure 15.7: PID controller



Provide details for FIC-1.

Figure 15.8: FIC



PID algorithm consists of three basic coefficients; proportional, integral, and derivative which are varied to get an optimal response. Before the working of the PID controller takes place, it should be tuned to suit with dynamics of the process to be controlled. Designers give the default values for P, I, and D terms, and these values couldn't give the desired performance and sometimes leads to instability and slow control performances.

Term P (Kc) is proportional to the current value of the SP − PV error. A high proportional gain results in a large change in the output for a given change in the error. If the proportional gain is too high, the system can become unstable. In contrast, a small gain results in a small output response to a large input error, and a less responsive or less sensitive controller.

Term I (Ti) accounts for past values of the SP − PV error and integrates them overtime to produce the I term. The integral term accelerates the movement of the process towards set-point and eliminates the residual steady-state error that occurs with a pure proportional controller. However, since the integral term responds to accumulated errors from the past, It can cause the present value to overshoot the set-point value.

Term D (Td) is the best estimate of the future trend of the SP − PV error, based on its current rate of change. Derivative action predicts system behavior and thus improves the settling time and stability of the system.

Table 15.1: Effect of increasing parameter independently

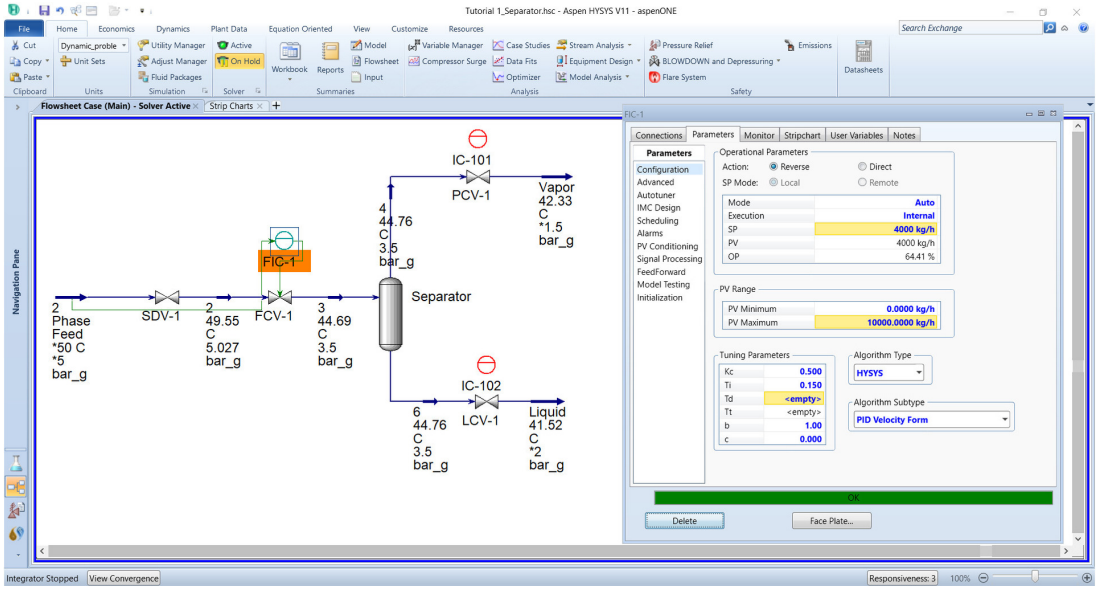
Param	Rise time	Overshoot	Settling time	Steady-state error	Stability
Kc	Decrease	Increase	Minor change	Decrease	Degrade
Ti	Decrease	Increase	Increase	Eliminate	Degrade
Td	Minor change	Decrease	Decrease	No effect	Improve

Use the below table to select tuning parameters.

Table 15.2: Tuning parameters

Control type	Gain	Integral time (min.)	Derivate time (min.)
Flow	0.4 – 0.65 (**0.5**)	0.05 – 0.25 (**0.15**)	–
Level	2 – 20 (**10**)	1 – 5 (**3**)	–
Pressure	2 – 10 (**5**)	2 – 10 (**5**)	–
Temperature	2 – 10 (**5**)	2 – 10 (**5**)	0 – 5 (**0**)

Figure 15.9: FIC



Provide details for PIC-1.

Figure 15.10: PIC

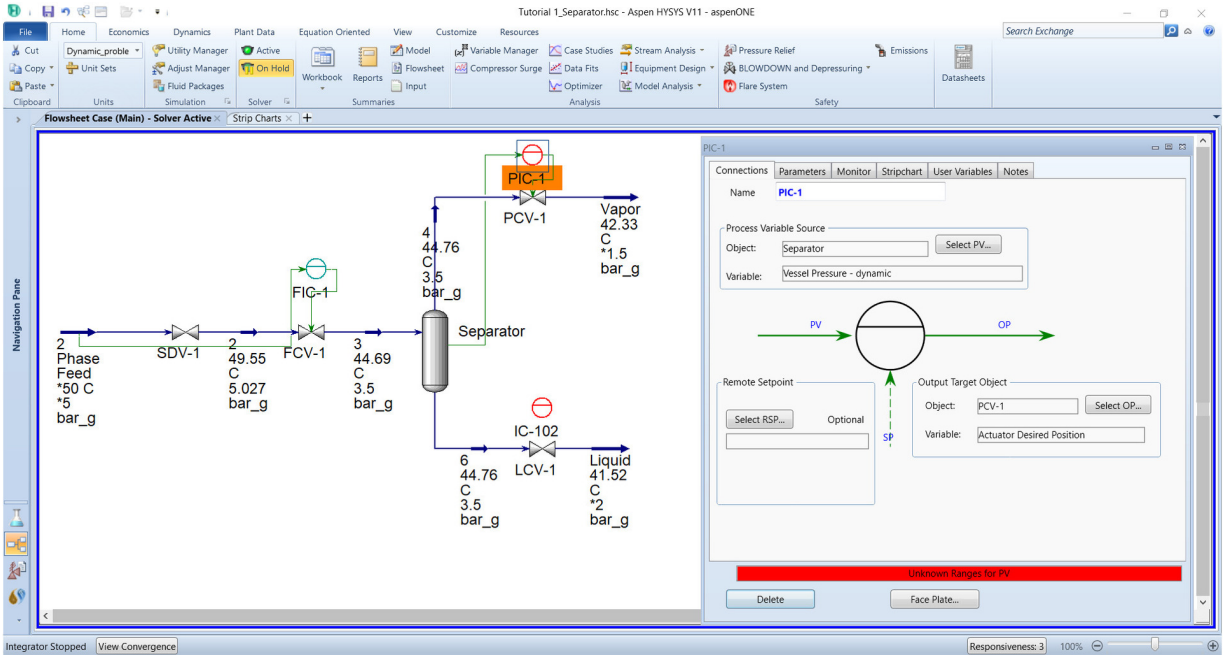
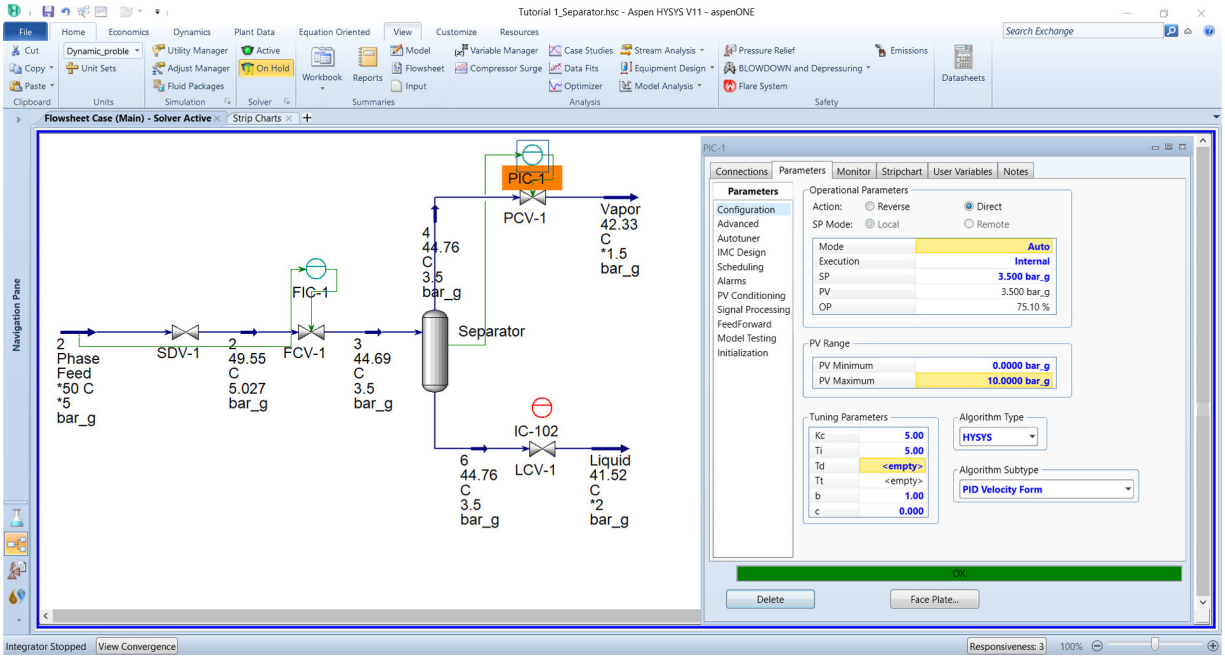


Figure 15.11: PIC



Provide details for LIC-1.

Figure 15.12: LIC

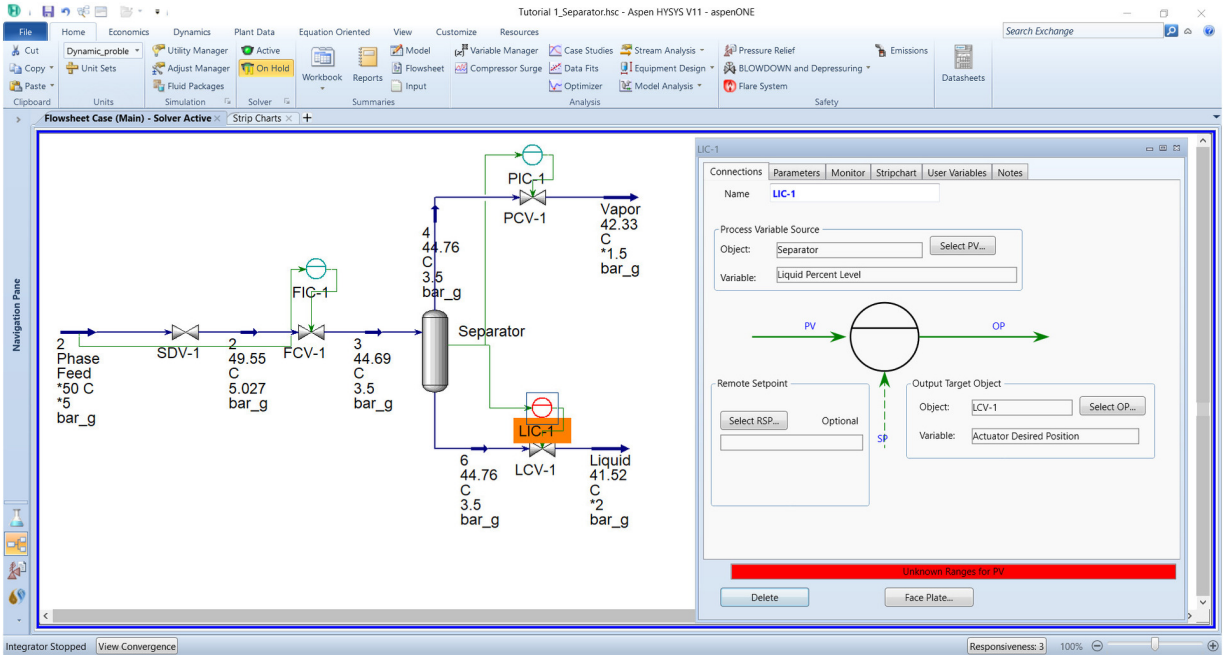
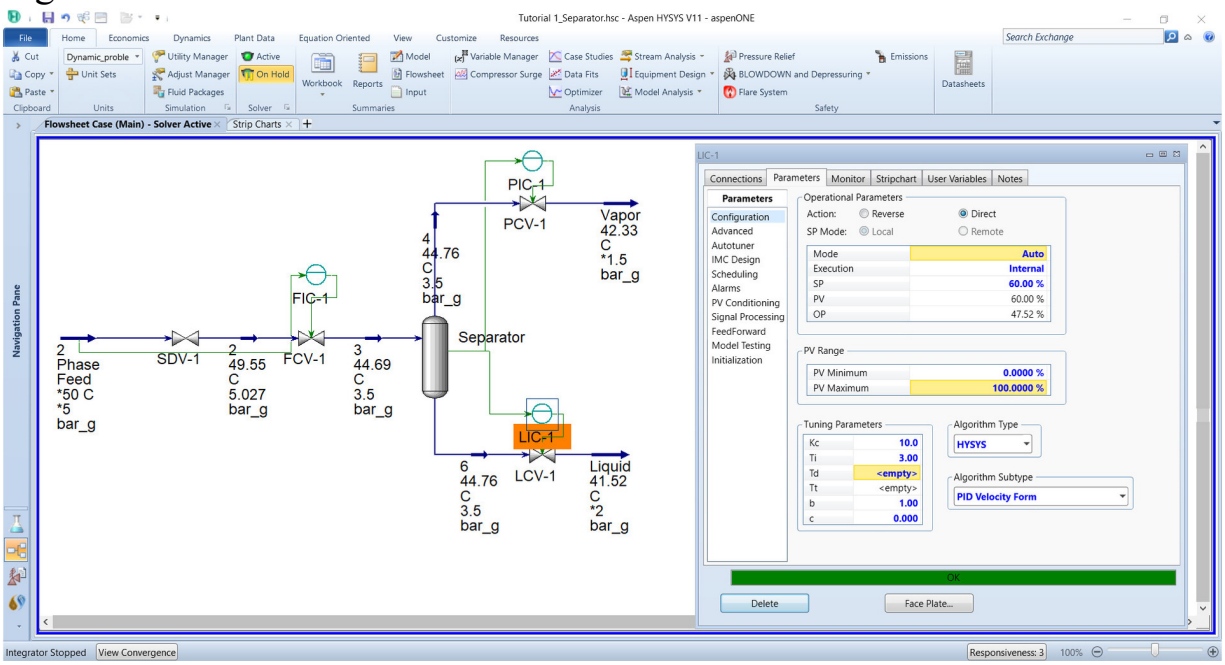


Figure 15.13: LIC



It is not possible to analyze the performance of a dynamic system without plotting different parameters w.r.to time. So we will provide a Strip chart for FIC-1.

Figure 15.14: Strip chart

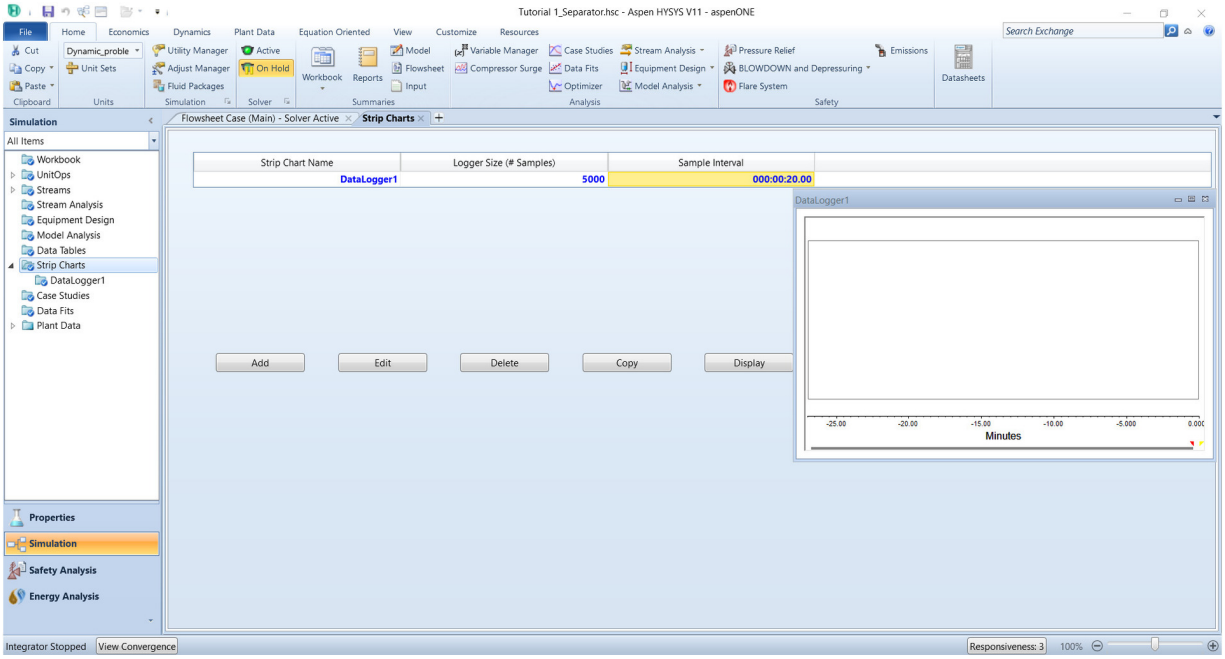
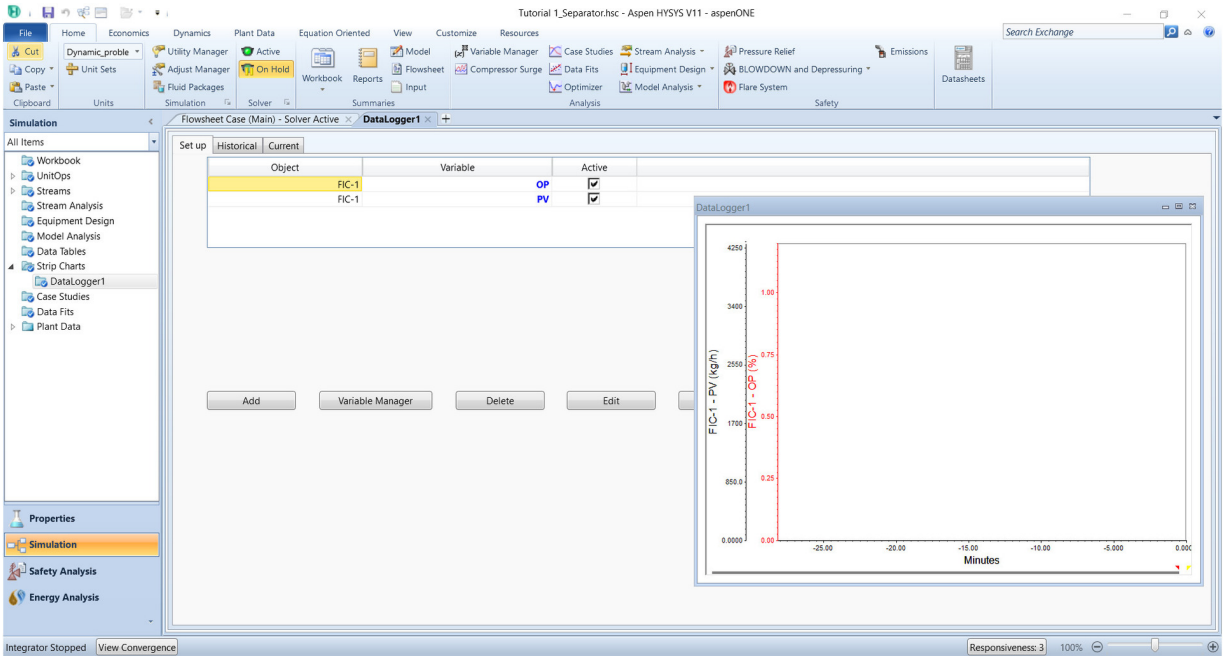
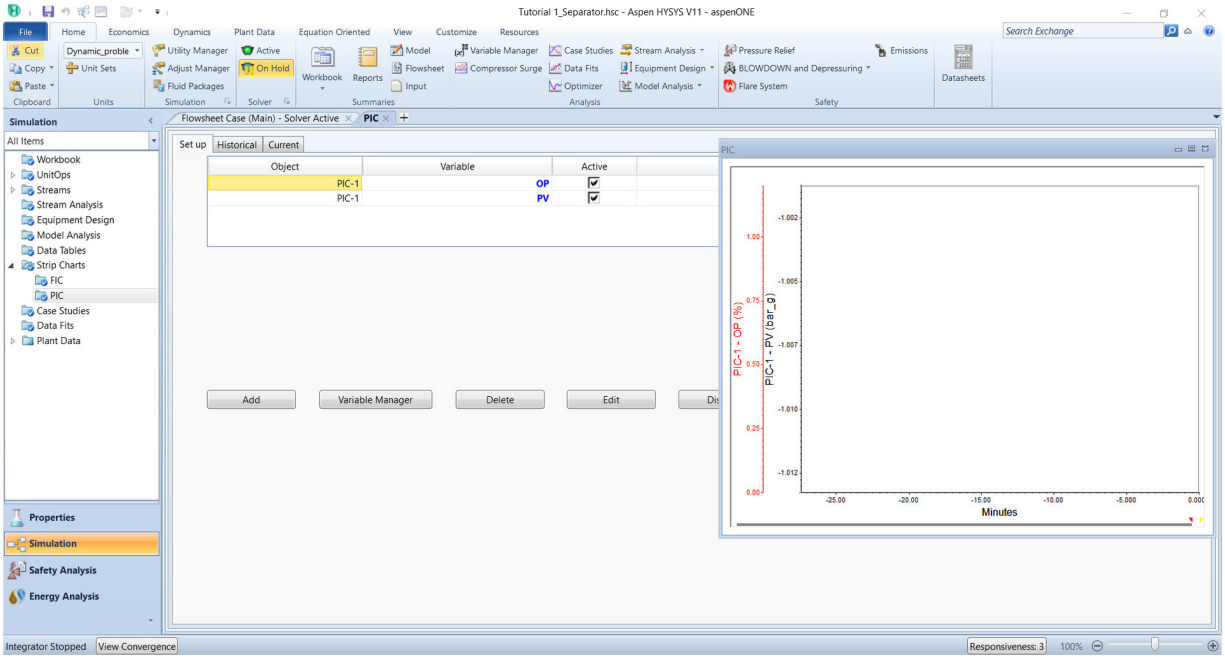


Figure 15.15: Strip chart



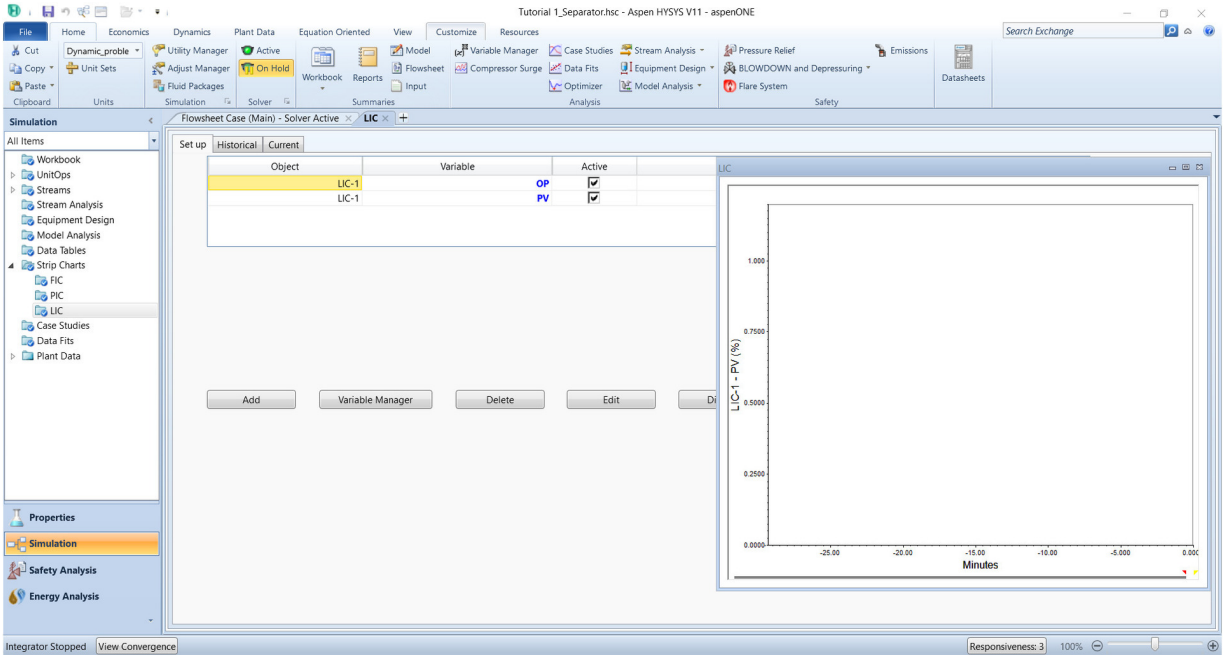
Provide a Strip chart for PIC-1.

Figure 15.16: Strip chart



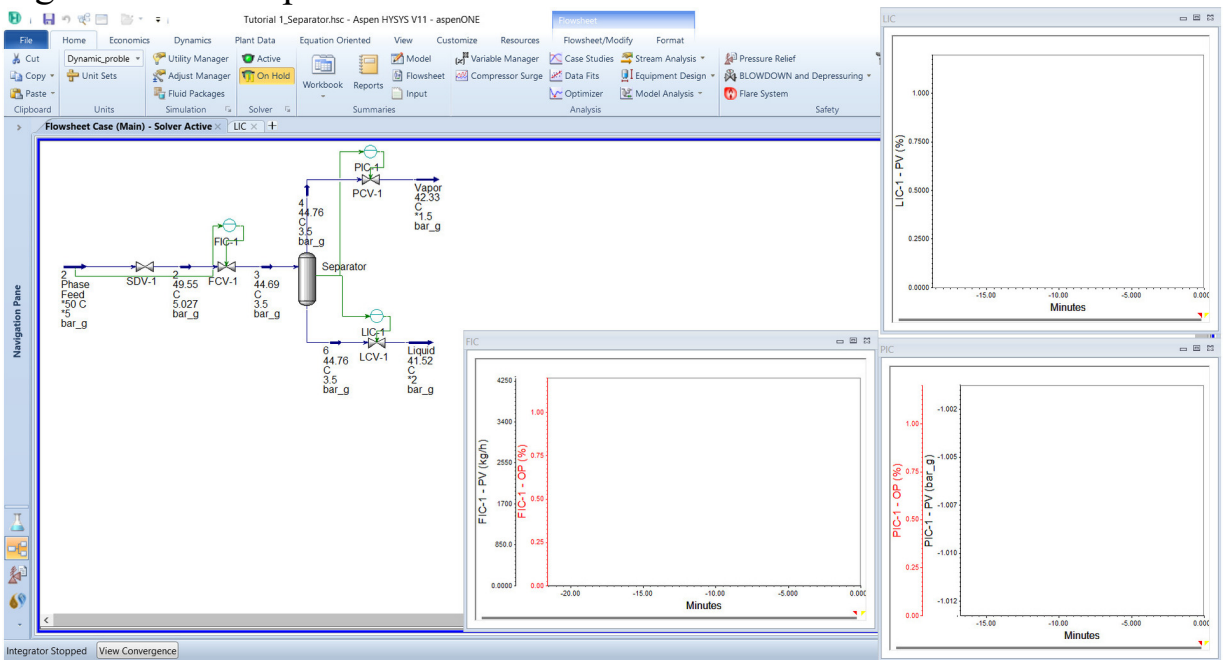
Provide Strip chart for LIC-1.

Figure 15.17: Strip chart



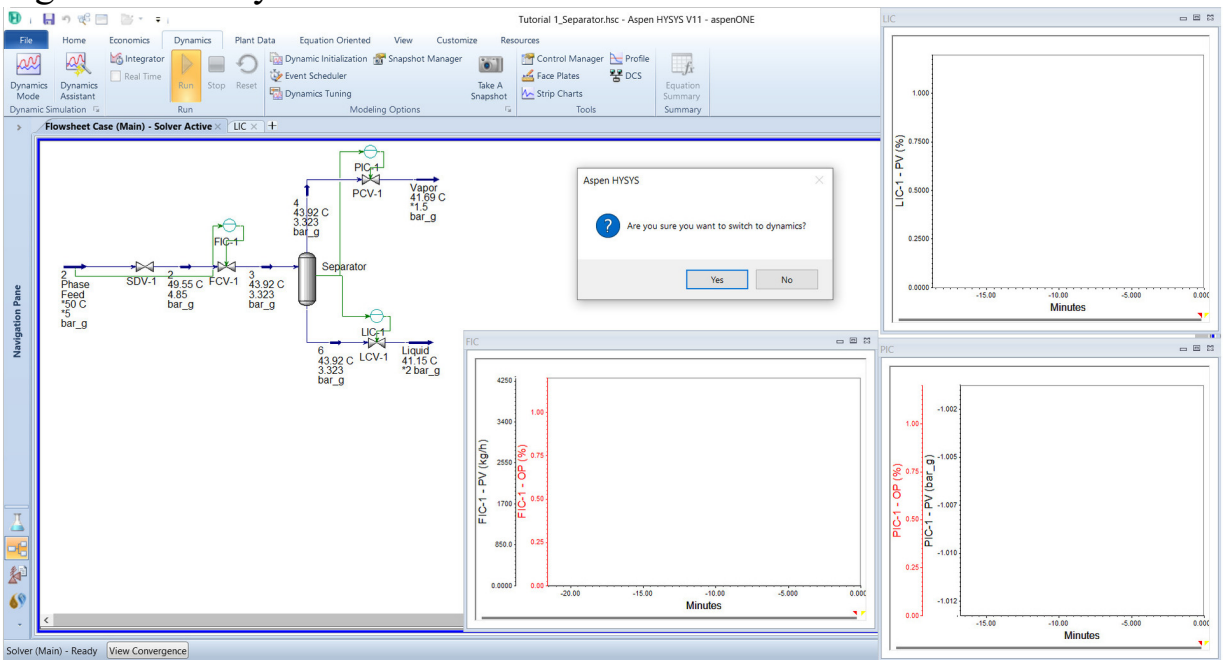
Provide Strip chart as shown below;

Figure 15.18: Strip chart



Click Dynamics\Dynamics Mode

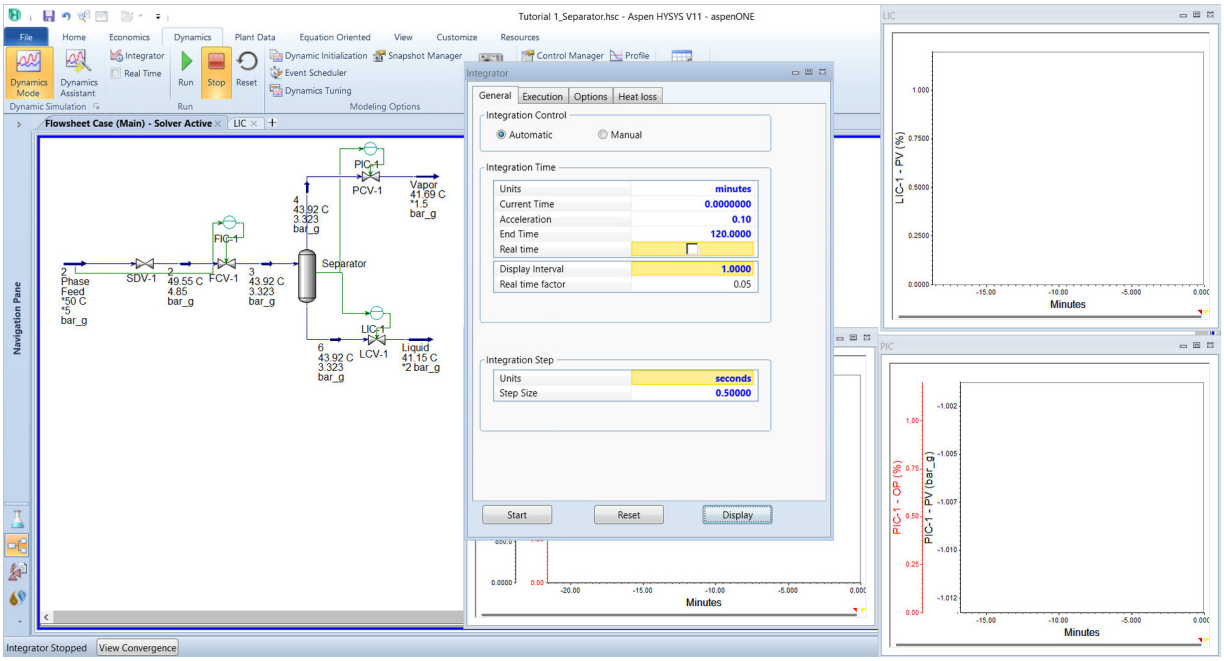
Figure 15.19: Dynamics



Click Dynamics\Integrator

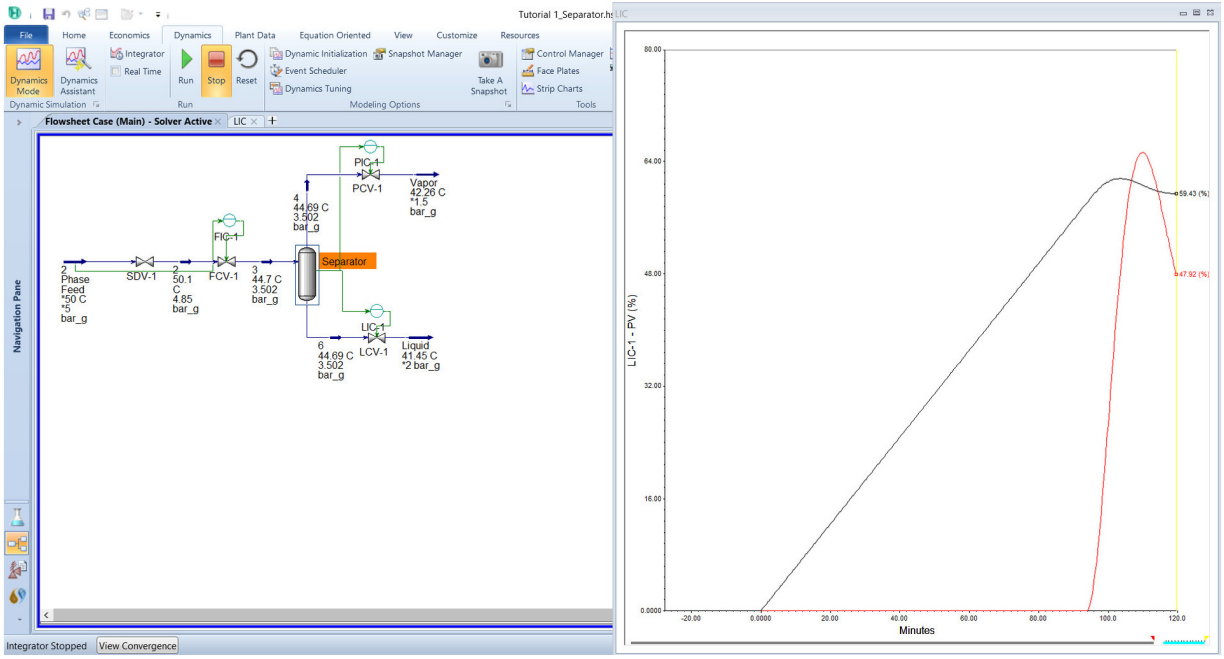
Start the simulation for 120 minutes. This is to check, how much time it takes to fill in the separator (14 m^3 * 60% = 8.4 m^3), when liquid in-flow is 5.4 m^3 /h.

Figure 15.20: Integrator 1



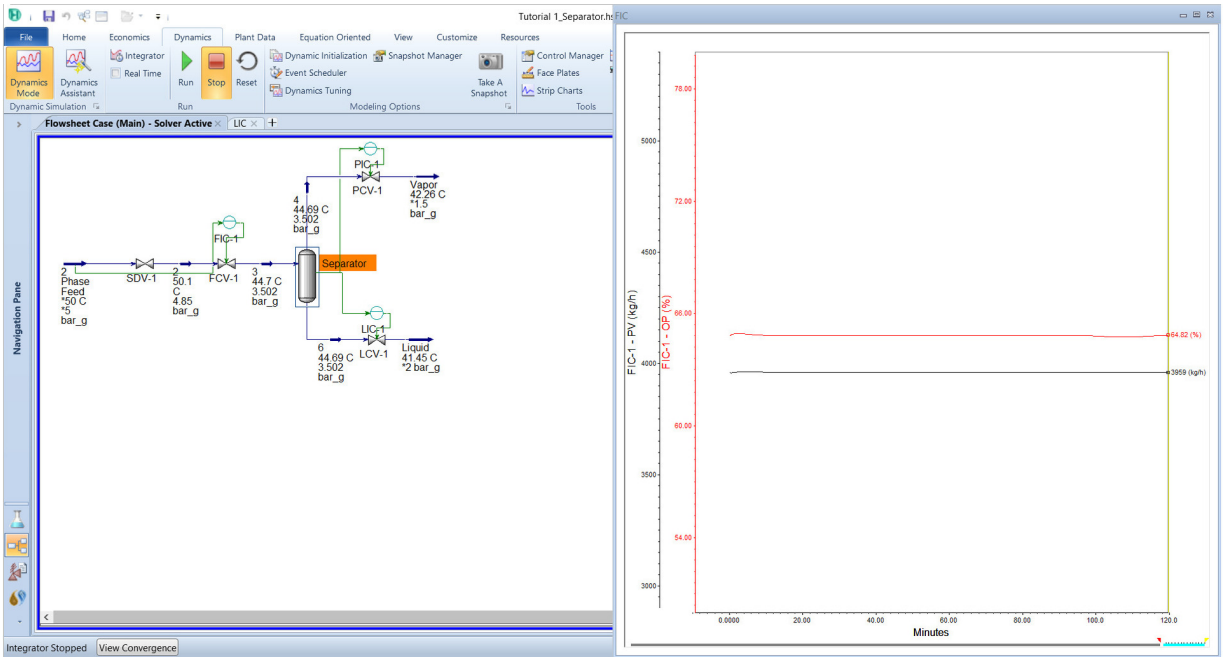
Refer below the plot (blue curve). It shows the separator level from 0% to 60%. It takes around 95 minutes to fill the separator to a 60% level.

Figure 15.21: Strip chart



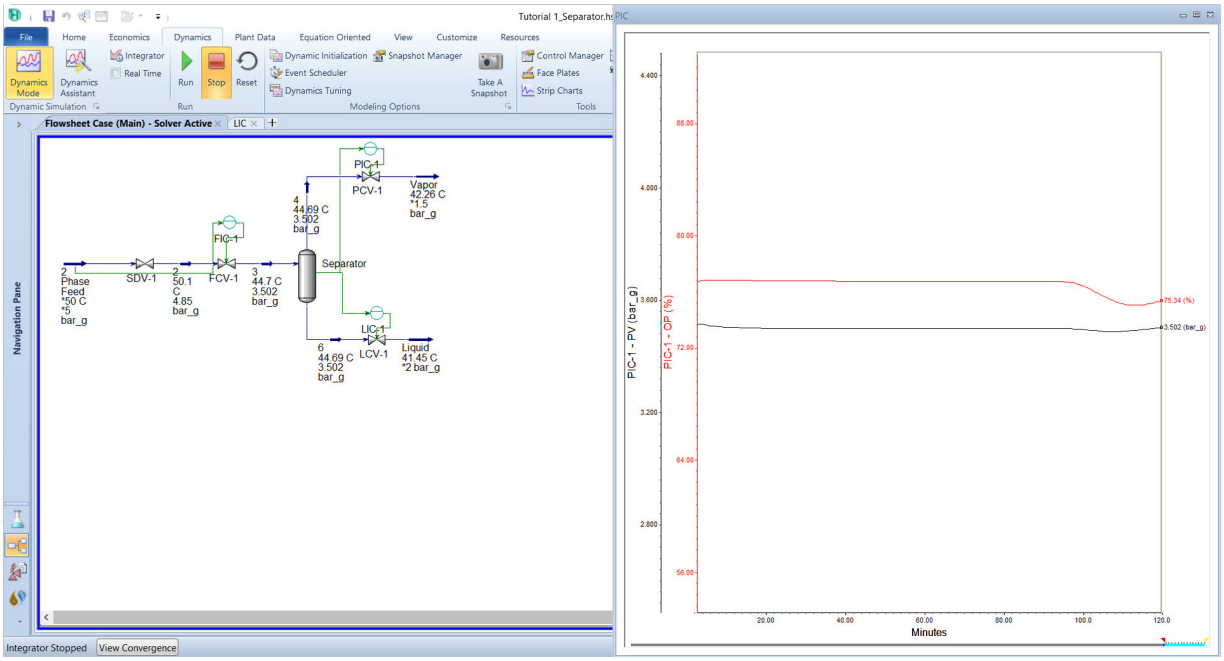
Similarly, refer below the plot for the FCV-1 valve opening (red curve) and FCV-1 flow rate. Valve opening is around 64%, which means the considered valve Cv and type is adequate.

Figure 15.22: Strip chart



Similarly, refer below the plot for the PCV-1 valve opening (red curve) and PCV-1 flow rate. Valve opening is around 43%, which means the considered valve Cv and type is adequate.

Figure 15.23: Strip chart



Let's try to simulate a case, where the pressure at inlet stream increases from 5 barg to 15 barg for 1 min and comes back to normal condition (5 barg) afterward.

Figure 15.24: Integrator

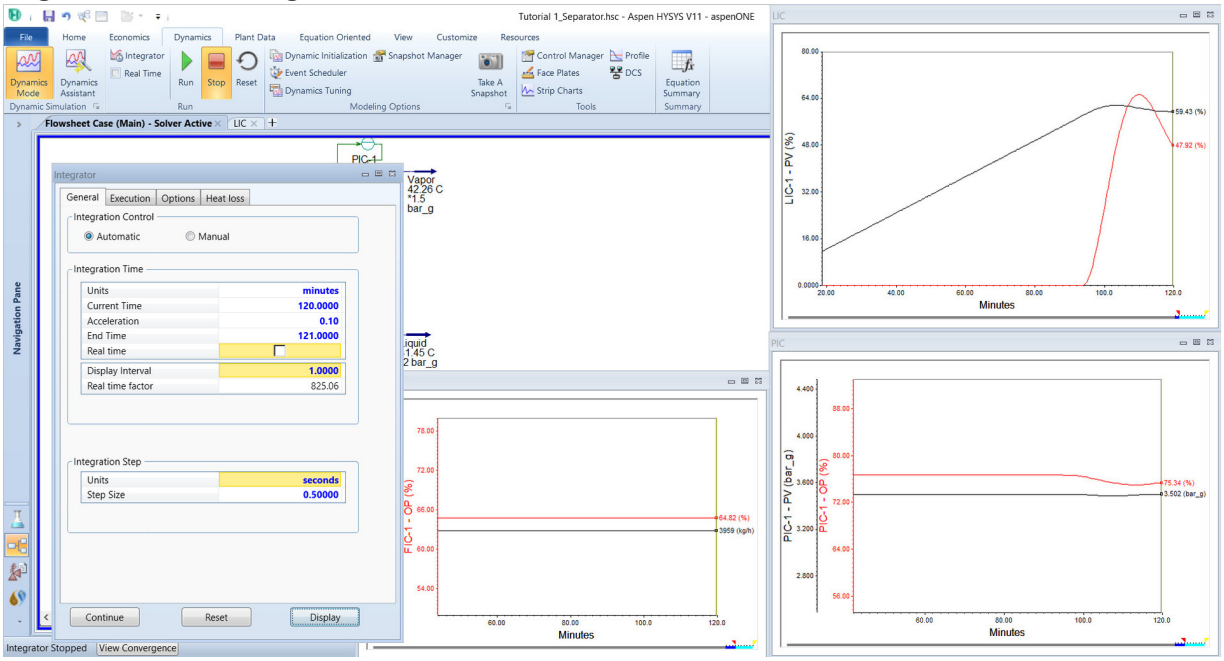
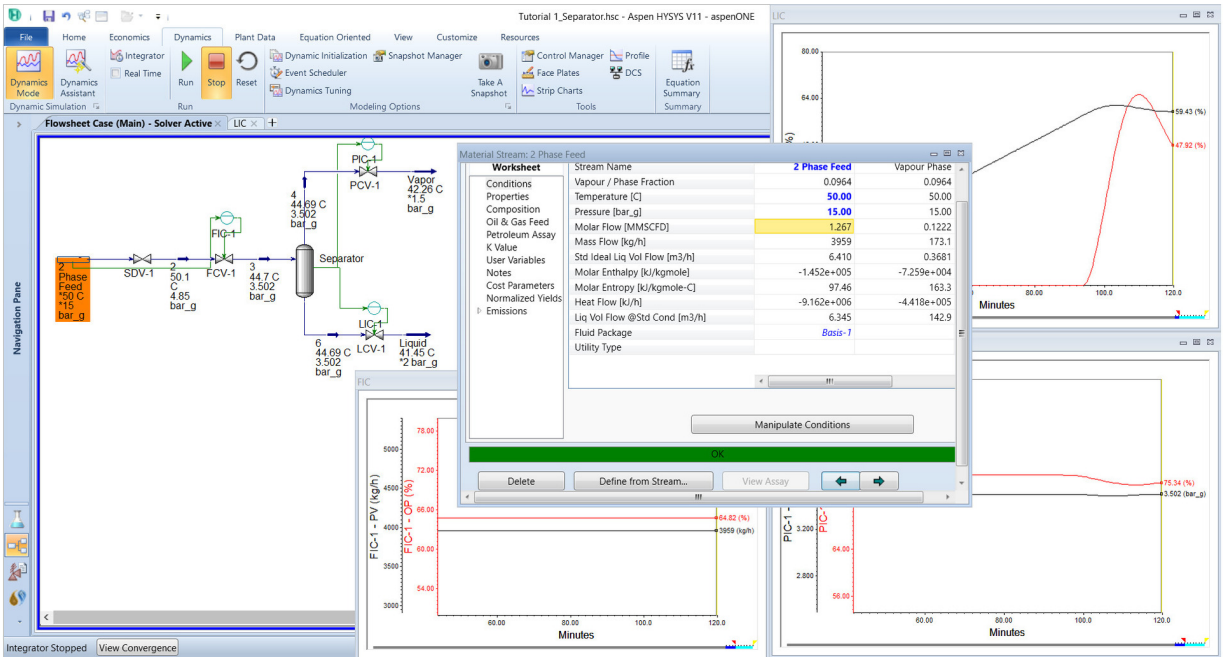


Figure 15.25: Material stream


Tutorial 1_Separator.hsc - Aspen HYSYS V11 - aspenONE
File
Home
Economics
Dynamics
Plant Data
Equation Oriented
View
Customize
Resources
Dynamics Mode
Dynamics Assistant
Integrator
Real Time
Run
Stop
Reset
Dynamic Initialization
Event Scheduler
Dynamics Tuning
Snapshot Manager
Take A Snapshot
Control Manager
Face Plates
Strip Charts
Profile
DCS
Equation Summary
Dynamic Simulation
Modeling Options
Tools
Summary
Flowsheet Case (Main) - Solver Active
LIC
Navigation Pane
PIC-1
PCV-1
Vapor
42.26 C
*1.5
bar_g
4
44.69 C
3.502
bar_g
FIC-1
2
Phase
Feed
*50 C
*15
bar_g
SDV-1
2
50.1
C
4.85
bar_g
FCV-1
3
44.7 C
3.502
bar_g
Separator
LIC-1
LCV-1
6
44.69 C
3.502
bar_g
Liquid
41.45 C
*2 bar_g
Material Stream: 2 Phase Feed
Worksheet
Conditions
Properties
Composition
Oil & Gas Feed
Petroleum Assay
K Value
User Variables
Notes
Cost Parameters
Normalized Yields
Emissions
Stream Name
2 Phase Feed
Vapour Phase
Vapour / Phase Fraction
0.0964
0.0964
Temperature [C]
50.00
50.00
Pressure [bar_g]
15.00
15.00
Molar Flow [MMSCFD]
1.267
0.1222
Mass Flow [kg/h]
3959
173.1
Std Ideal Liq Vol Flow [m3/h]
6.410
0.3681
Molar Enthalpy [kJ/kgmole]
-1.452e+005
-7.259e+004
Molar Entropy [kJ/kgmole-C]
97.46
163.3
Heat Flow [kJ/h]
-9.162e+006
-4.418e+005
Liq Vol Flow @Std Cond [m3/h]
6.345
142.9
Fluid Package
Basis-1
Utility Type
Manipulate Conditions
OK
Delete
Define from Stream...
View Assay
LIC
59.43 (%)
47.92 (%)
Minutes
FIC
FIC-1 - PV (kg/h)
FIC-1 - OP (%)
64.82 (%)
3959 (kg/h)
Minutes
PIC-1
PIC-1
75.34 (%)
3.502 (bar_g)
Minutes
Integrator Stopped
View Convergence

Run the integrator for 1 minute. Change the inlet stream pressure back to 5 barg.
Run the integrator for 360 minutes.

Figure 15.26: Material stream

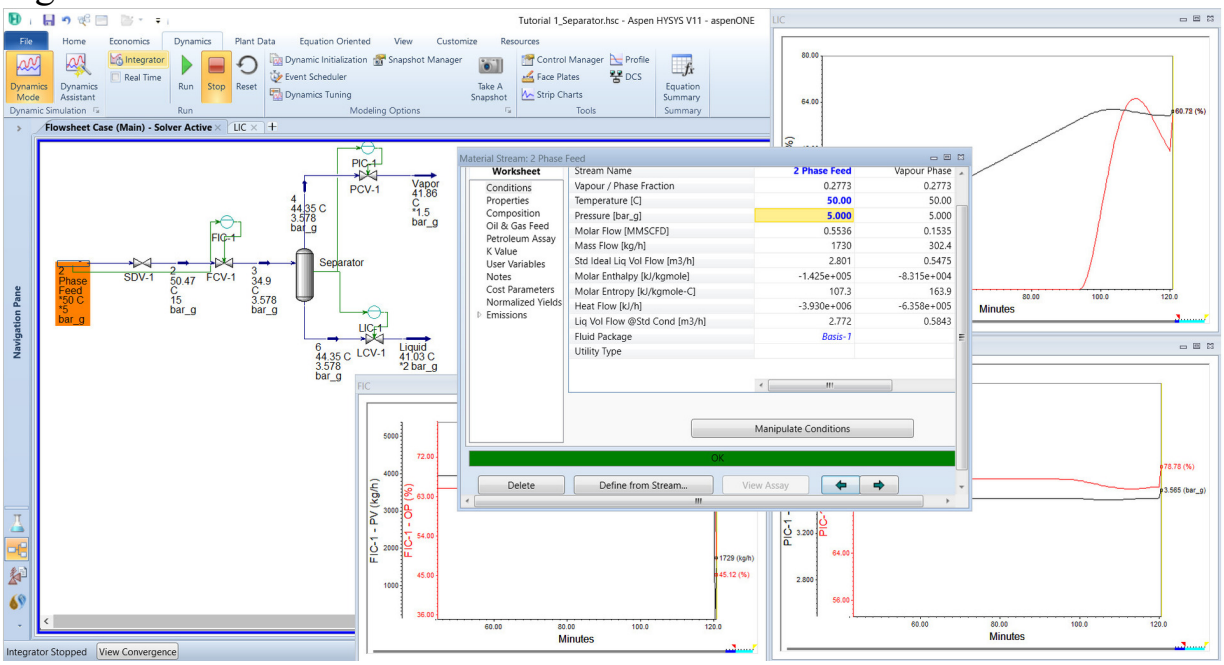
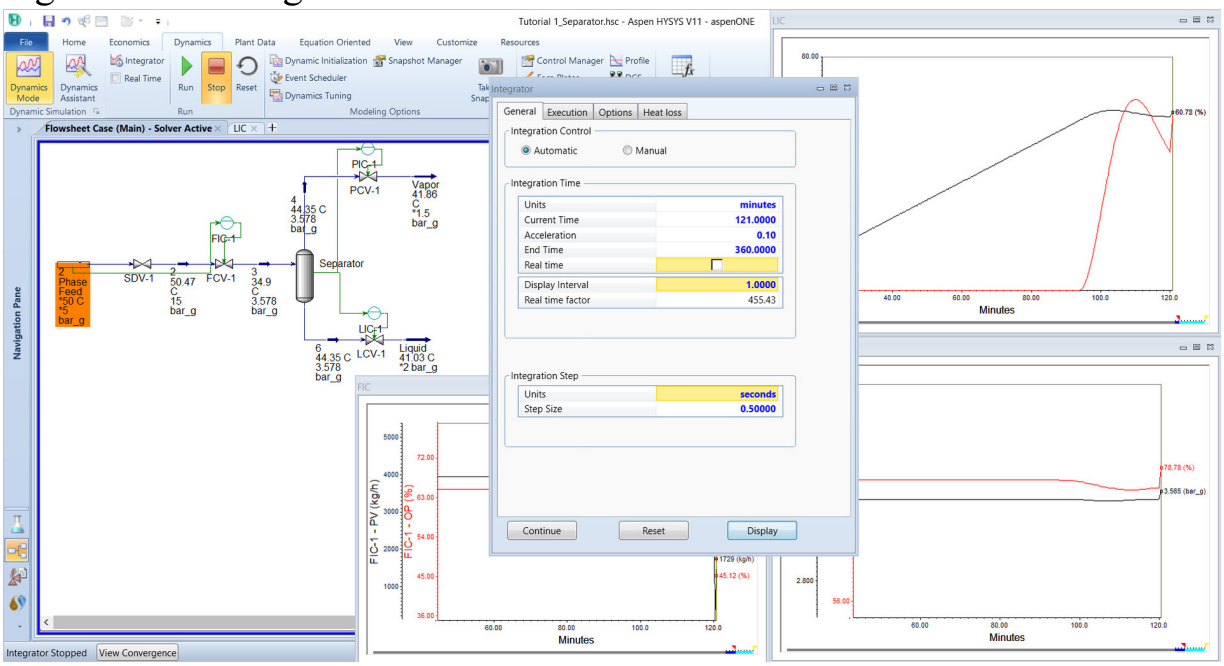
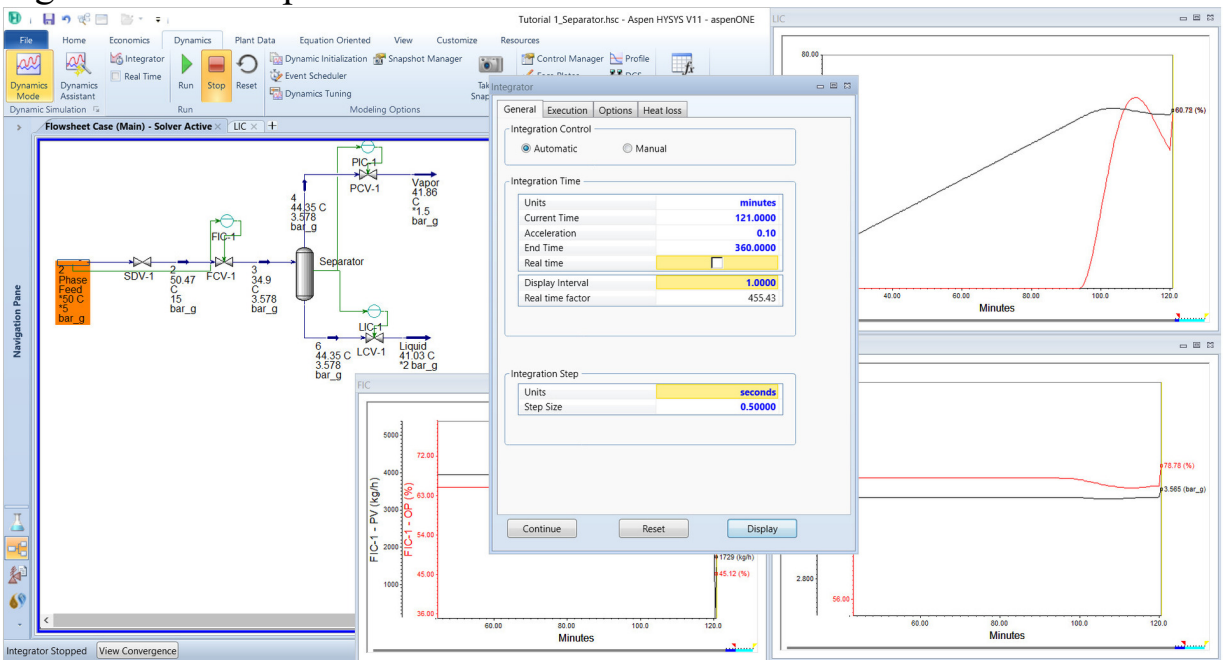


Figure 15.27: Integrator



As expected, FCV-1 shall get throttled to maintain the set flow rate.

Figure 15.28: Strip chart



There doesn't seem to be much impact on PIC-1 or LIC-1 functioning as well.

Figure 15.29: Strip chart

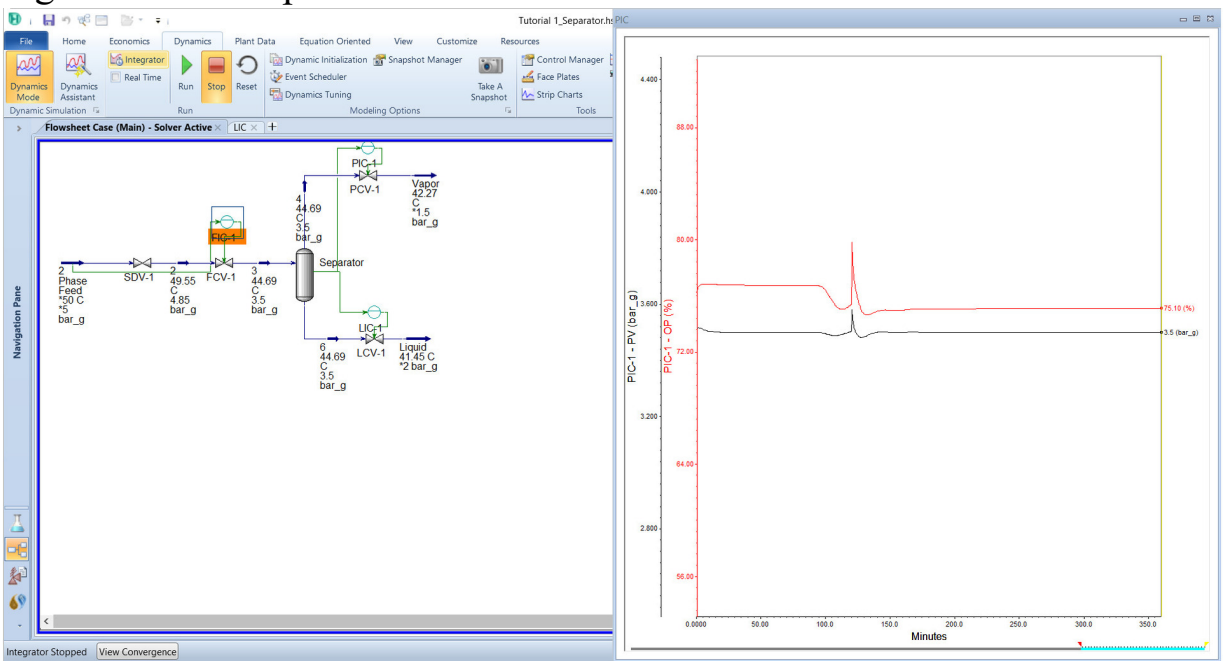
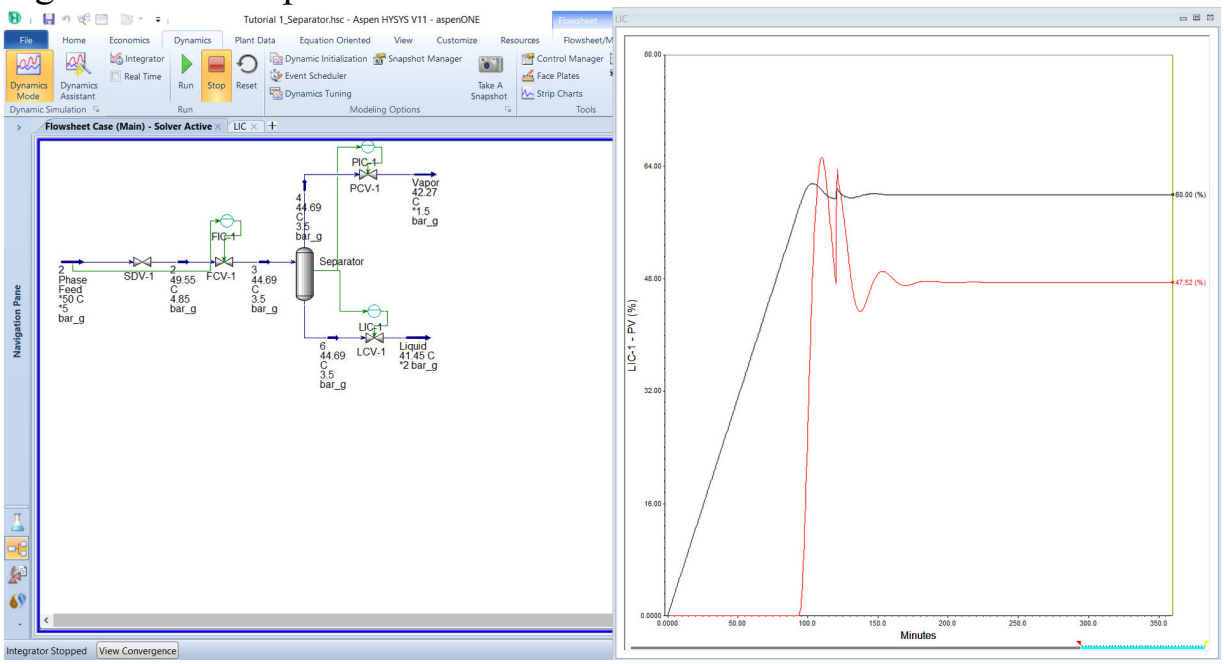


Figure 15.30: Strip chart



Let's prepare a cause and effect matrix. When LIC-1 goes High, then it should give an alarm. And when LIC-1 goes High High, then it should give an alarm, and SDV-1 shall close.

Figure 15.31: Digital control point

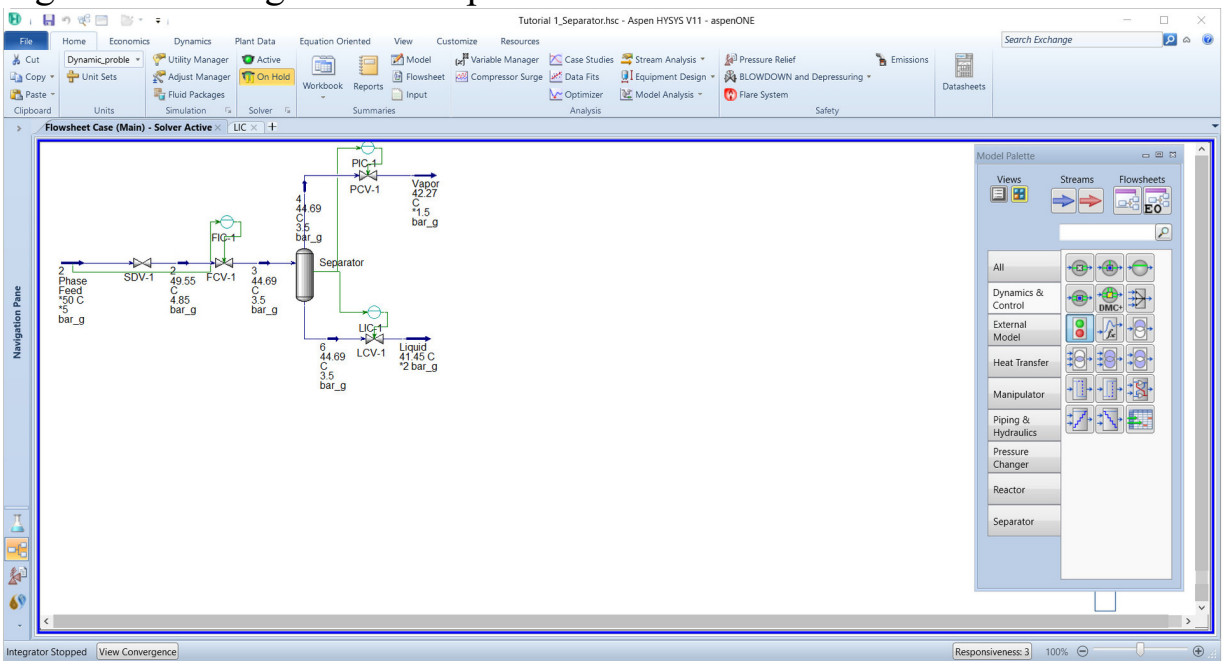


Figure 15.32: Digital control point

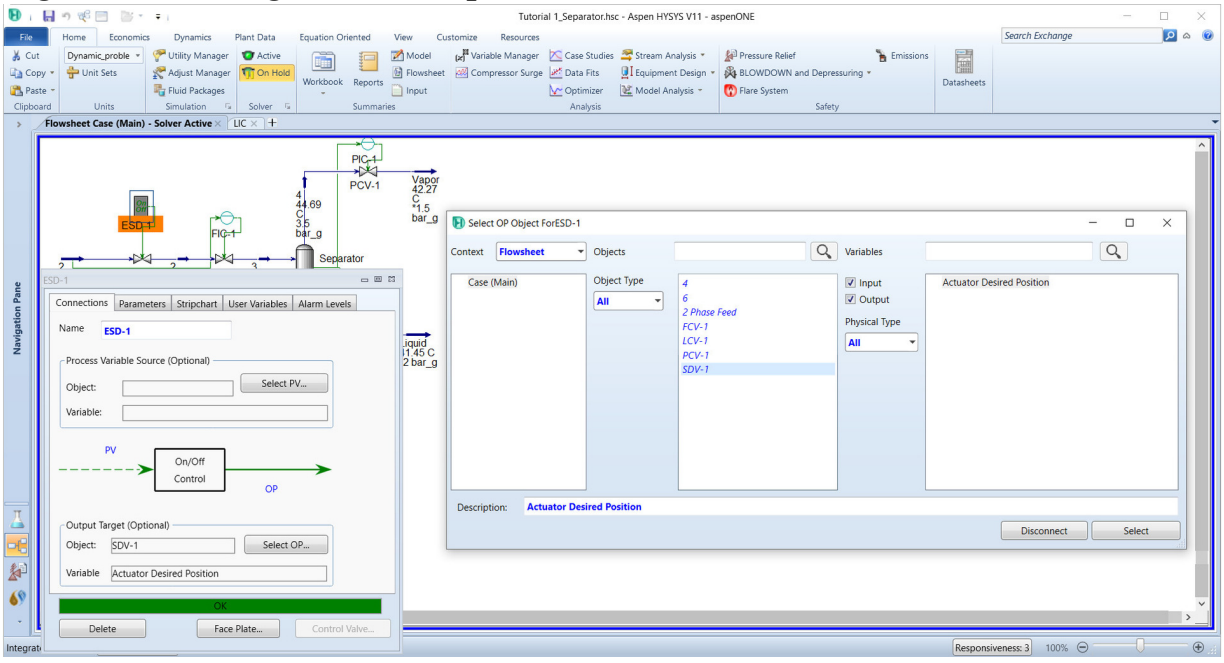
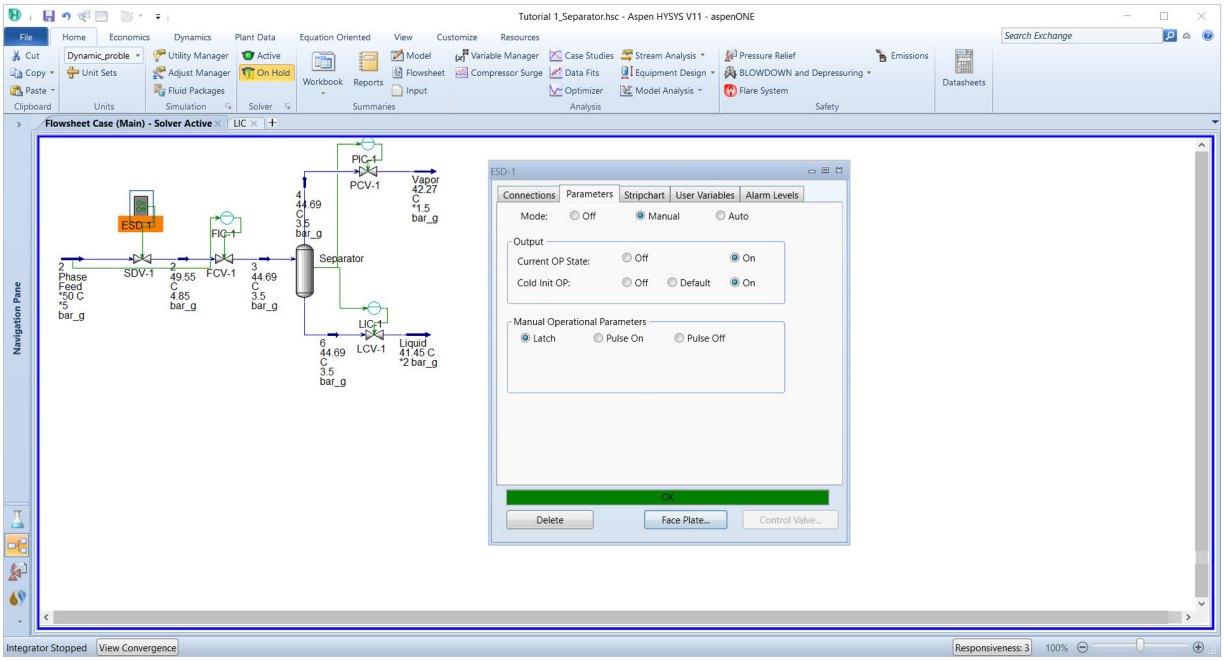


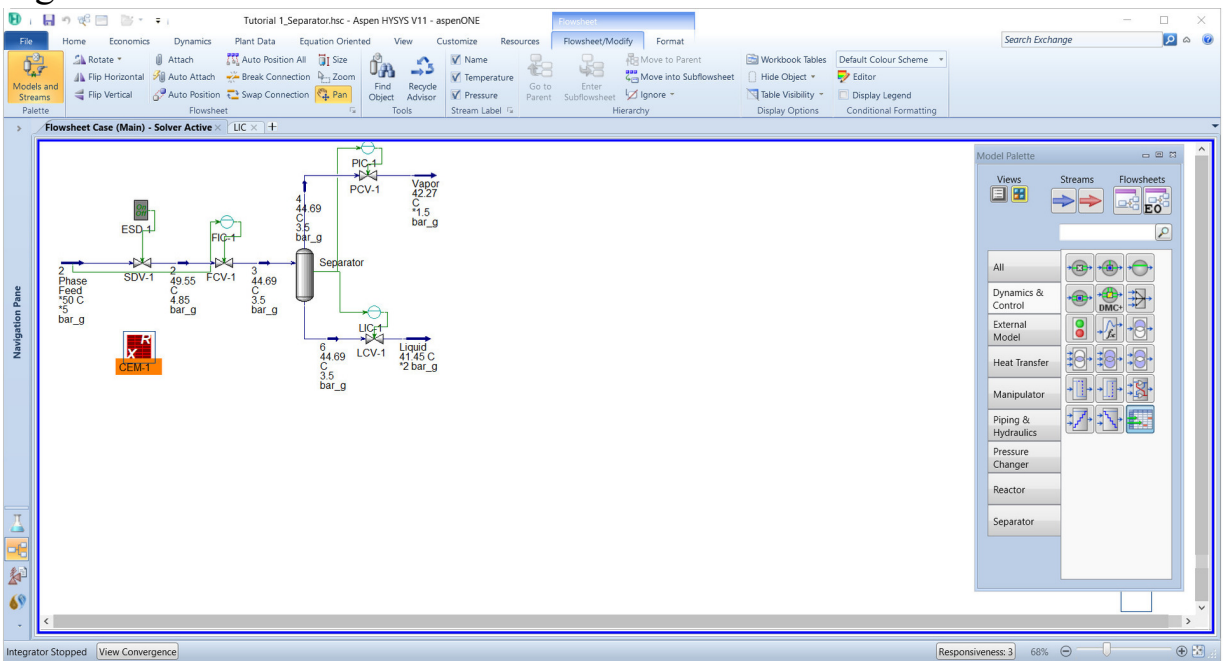
Figure 15.33: Digital control point



A cause and effect diagram examines why may happen by organizing potential causes into smaller categories. It can also be useful for showing relationships between contributing factors. It is often referred to as a fishbone diagram or Ishikawa diagram.

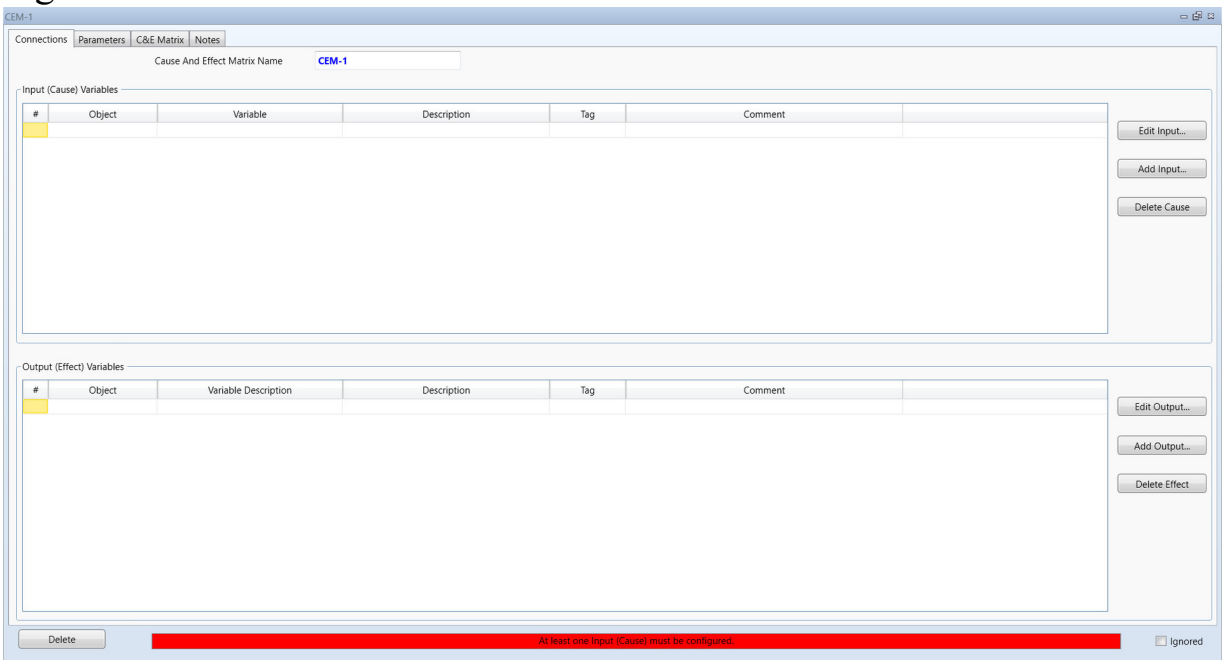
Let's add the Cause & Effect Matrix for our system. Select "Cause-and-Effect Matrix" available in Model Palette.

Figure 15.34: Cause & effect matrix



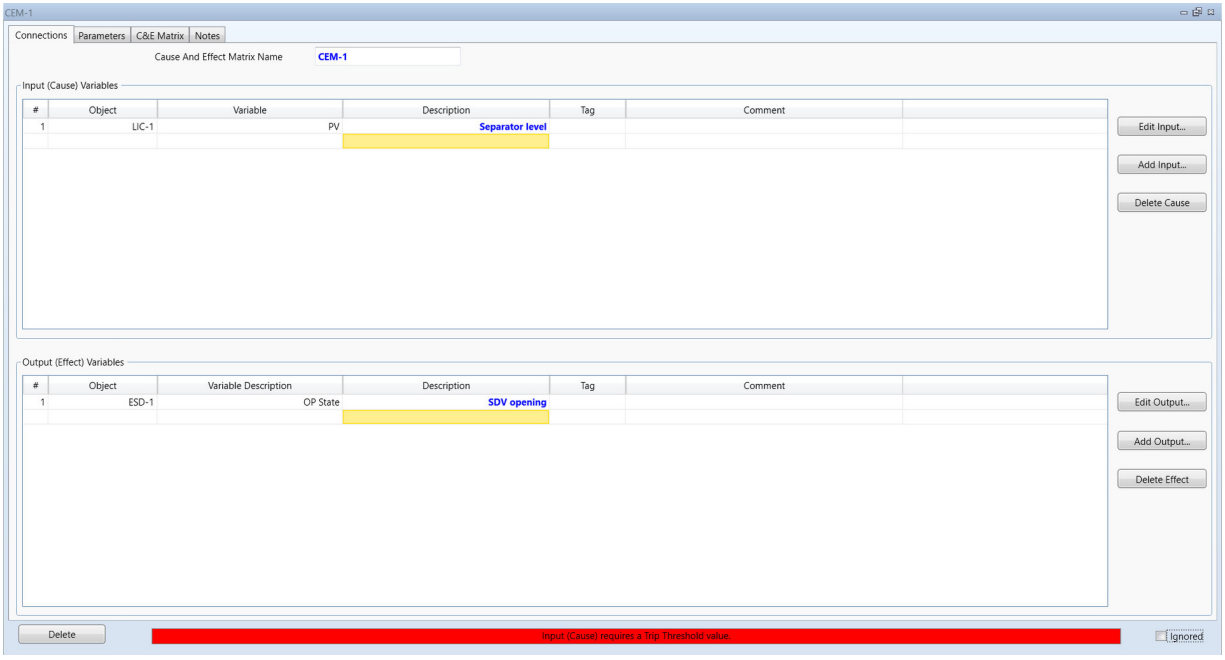
Double-clicking on CEM-1 shall open a window.

Figure 15.35: Cause & effect matrix



"Add Input" in "Input (Cause) Variables" and "Add Output" in "Output (Effect) Variables" as shown below.

Figure 15.36: Cause & effect matrix



Separator level Alarm will sound at 70% level and Effect (SDV-1 closure) will initiate at 80% level. In this case "High" checkbox is checked, as trip

happens at High High condition. If there was any trip action for Low Low condition, then this box would be kept unchecked.

Figure 15.37: Cause & effect matrix

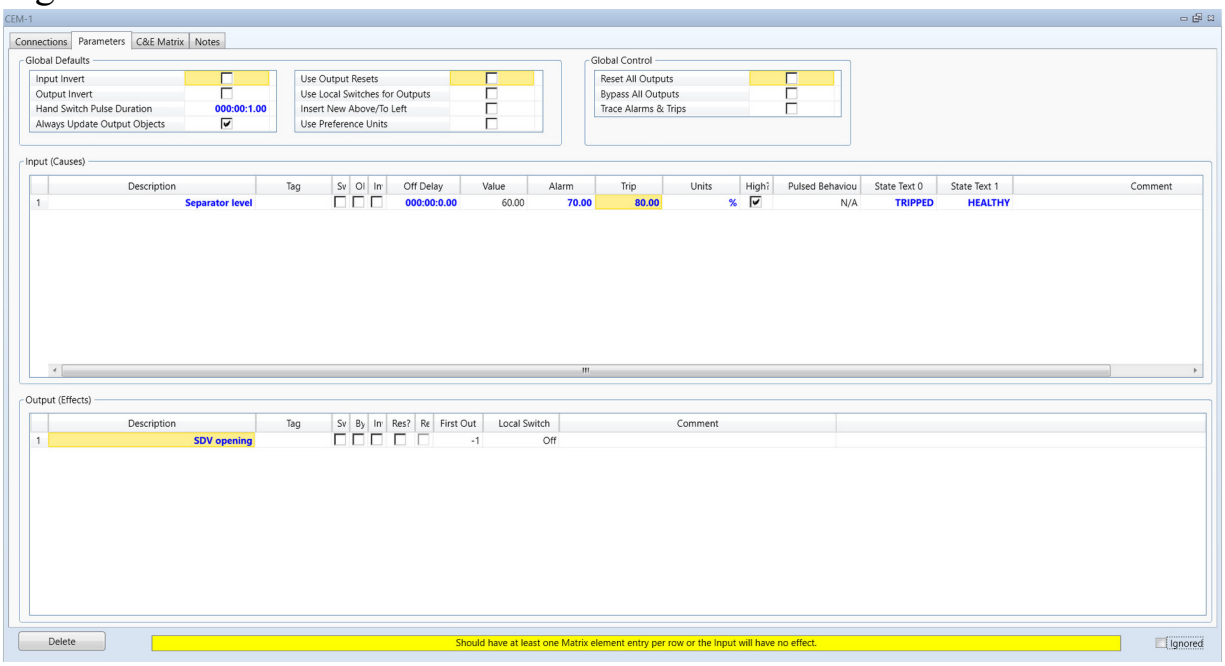
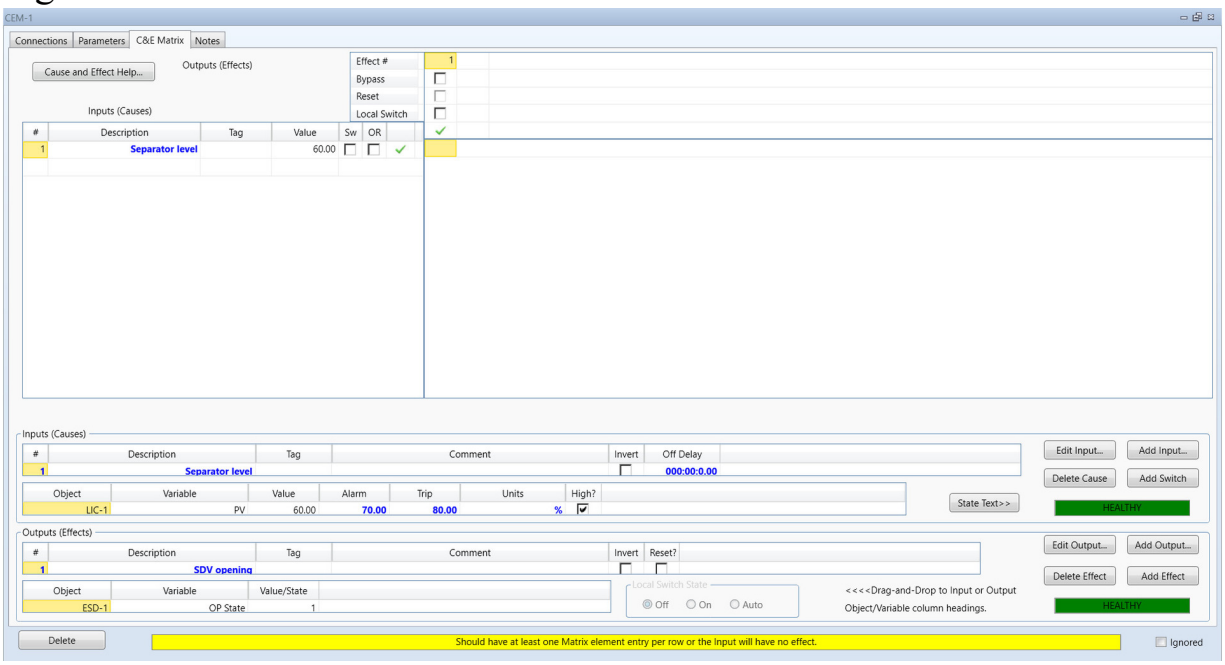
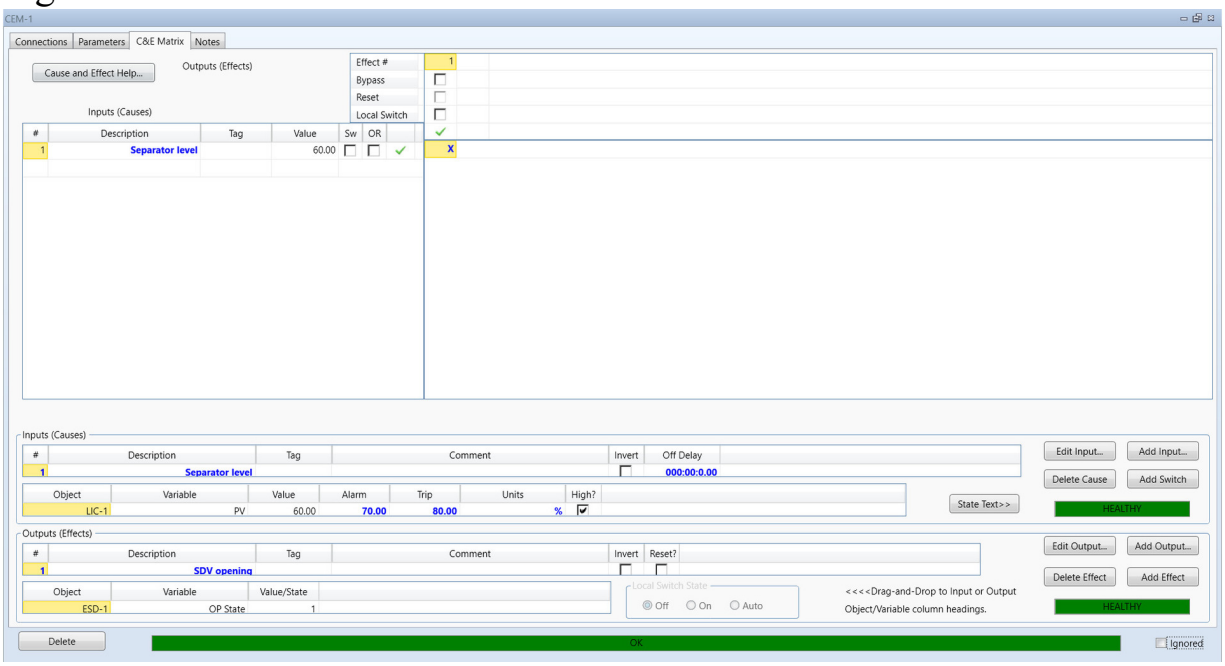


Figure 15.38: Cause & effect matrix



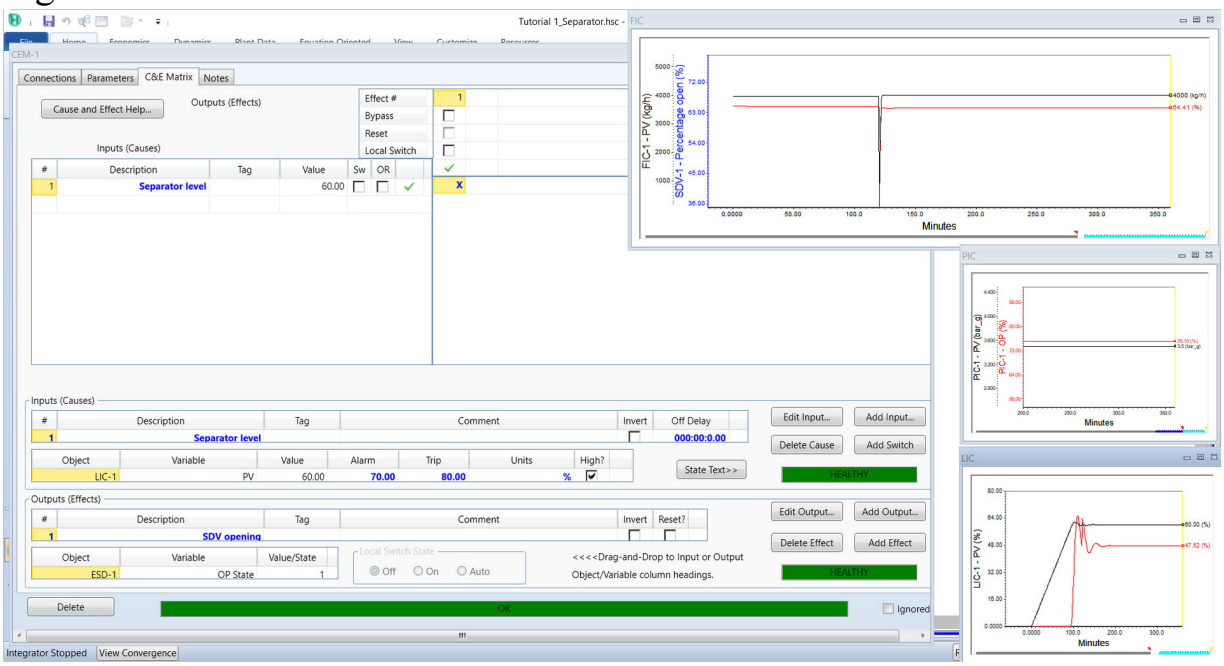
There may be several different Causes and their Effects on the plant. Whatever relevant Causes and Effects shall be marked “X” as shown below.

Figure 15.39: Cause & effect matrix



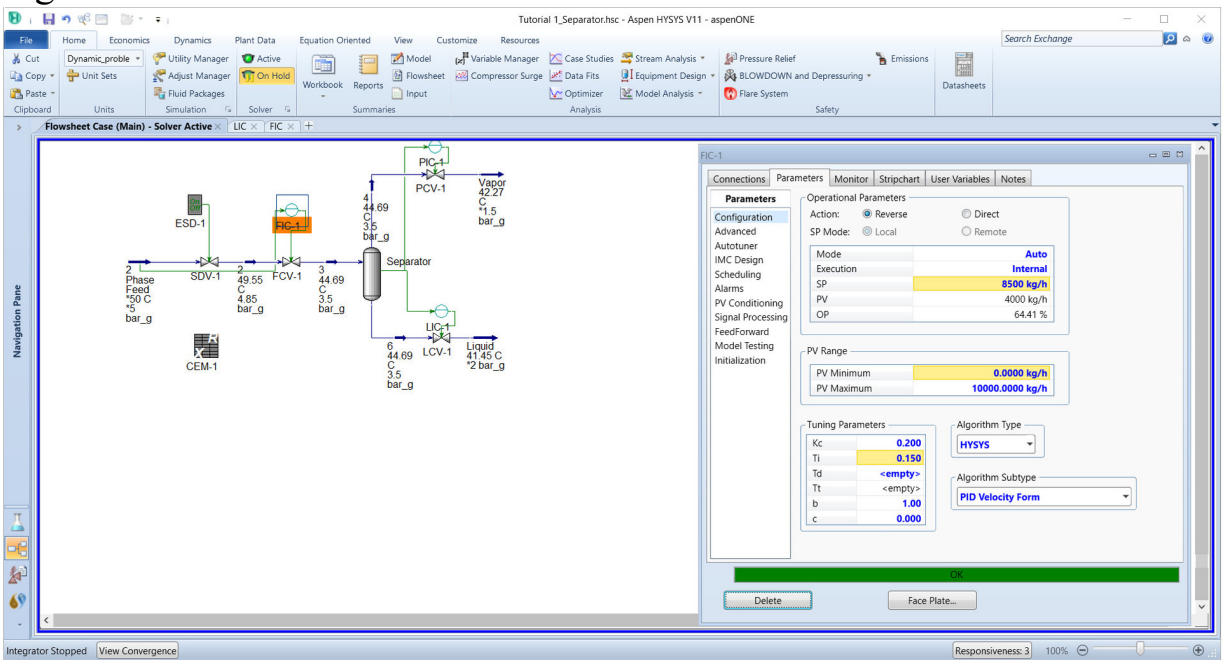
Add SDV-1 valve opening parameter in the FCV plot.

Figure 15.40: Cause & effect matrix



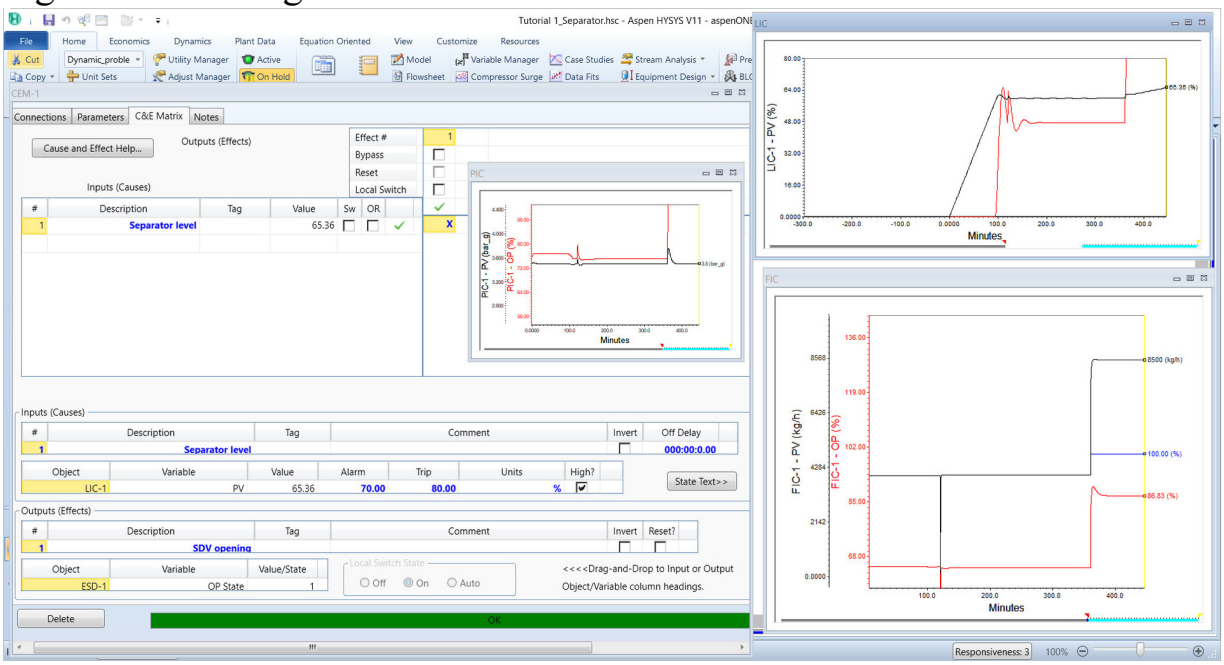
Let’s increase the FIC-1 set point from 4000 kg/h to 8500 kg/h. It shall increase the separator level. Change Kc of FIC-1 to 0.2.

Figure 15.41: FIC



Increase integrator end time to 1000 minutes.

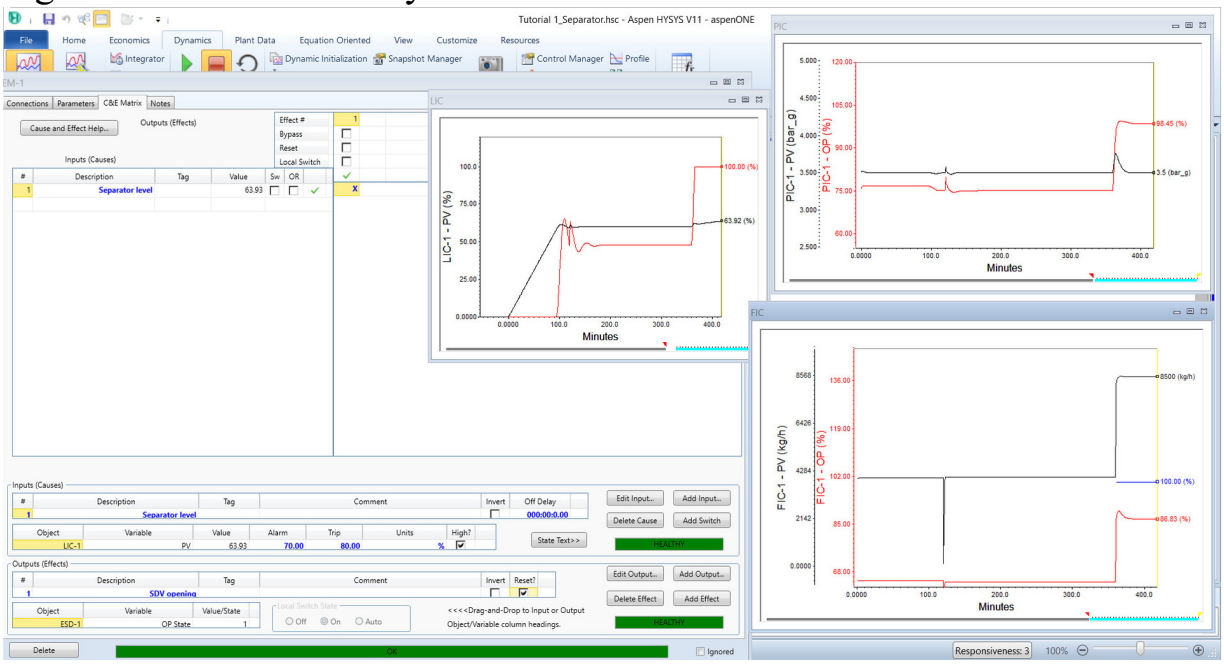
Figure 15.42: Integrator



Start the Integrator.

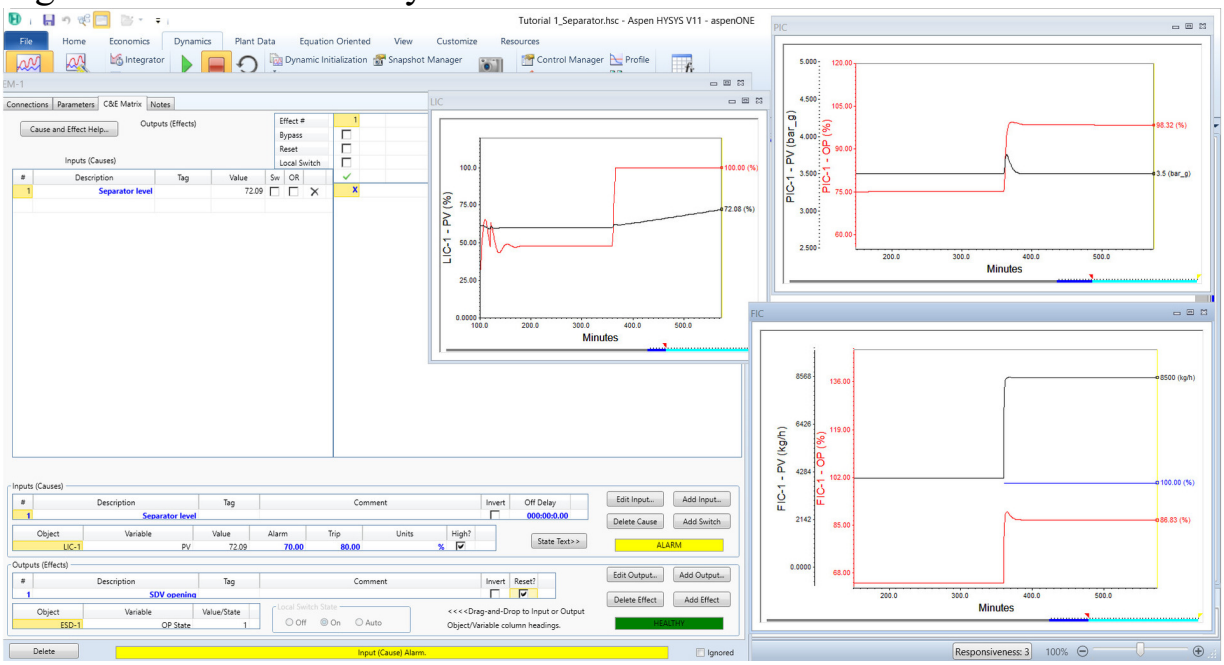
Soon flow rate increases to 8500 kg/h. LIC-1 and PIC-1 OP reaches to 100%, as LCV-1 and PCV-1 CVs are not adequate for this high flow rate. Separator level starts to increase.

Figure 15.43: Model analysis



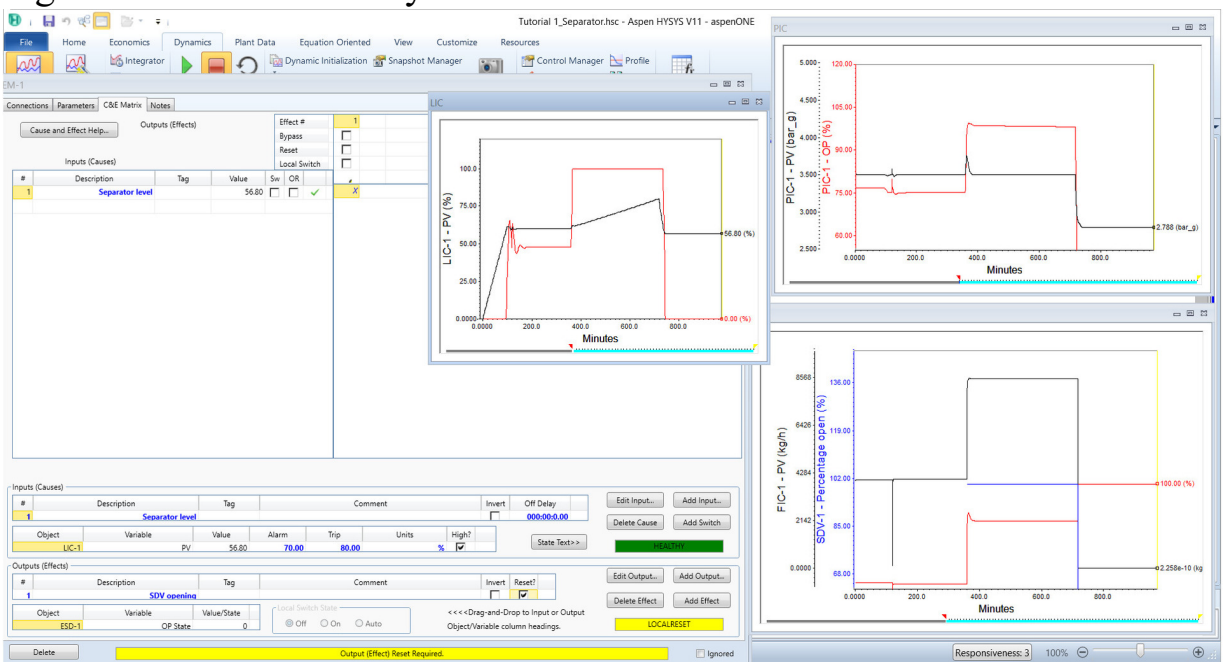
C&E Matrix shows Alarm (yellow bar), when the separator level reaches 70% (High Alarm).

Figure 15.44: Model analysis



Further running of the integrator shall result in separator level rise exceeding 80% and closure of SDV-1. This results in mass flow rate dropping to zero.

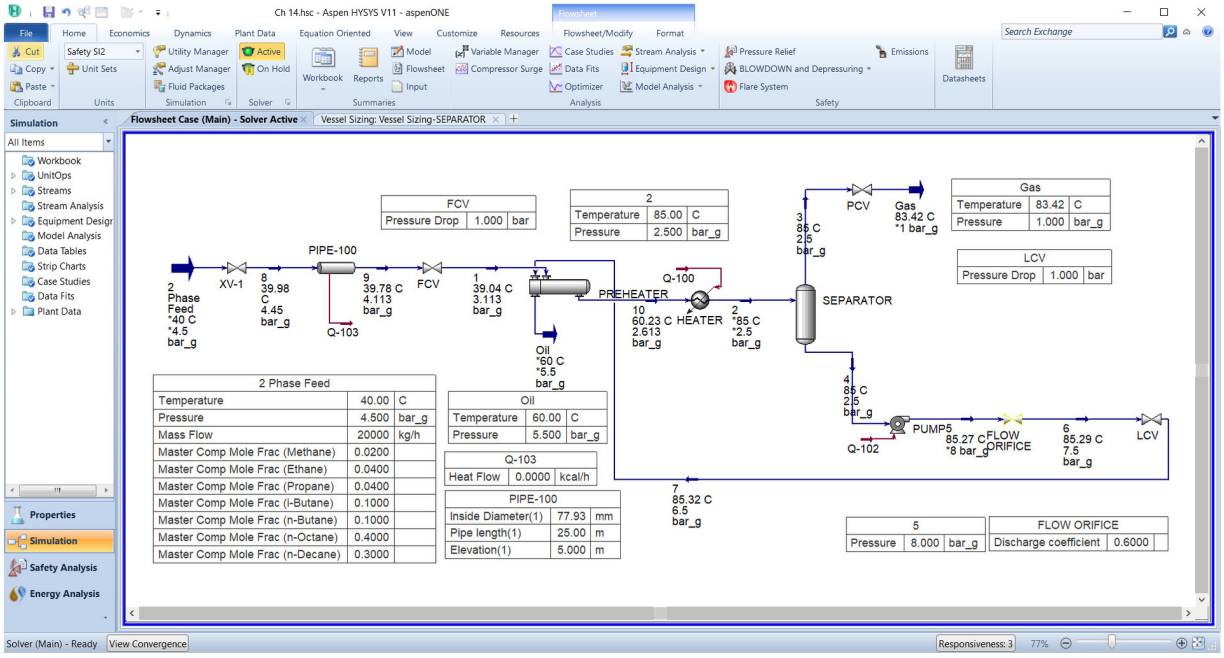
Figure 15.45: Model analysis



CHAPTER 16: ASPEN HYSYS DYNAMICS (PROJECT PROBLEM)

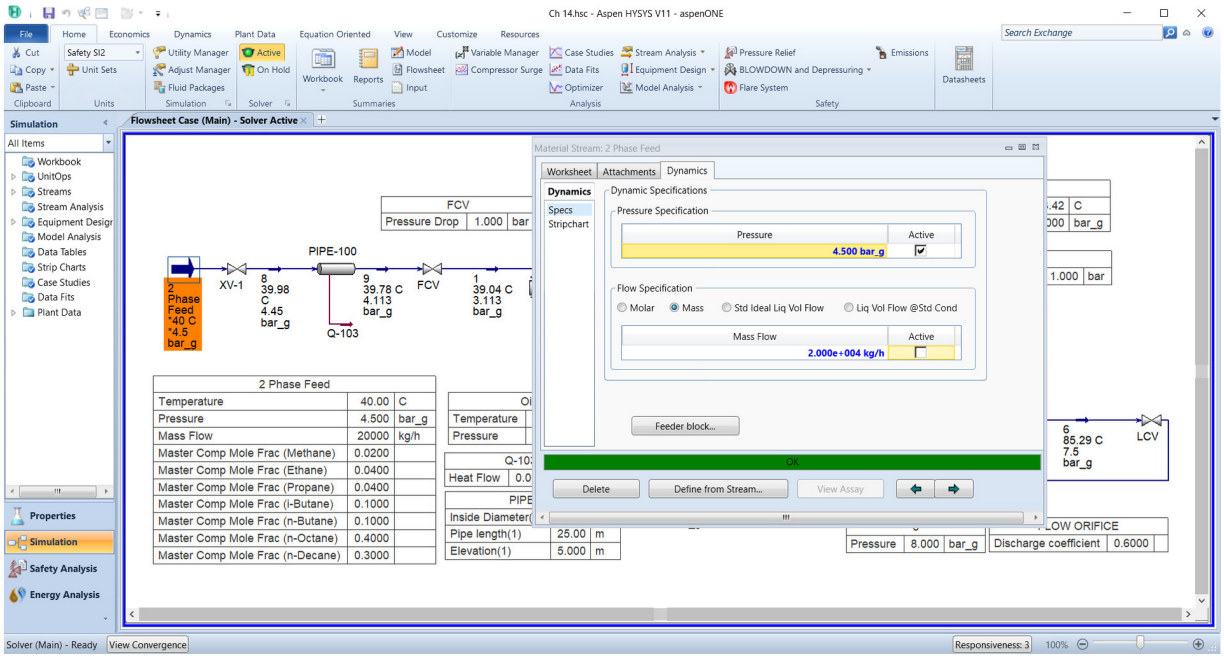
We have already designed a typical oil and gas separation facility in chapter 12.
In this chapter, we are going to convert that steady-state model into a dynamic model.

Figure 16.1: Model



Make sure the Aspen HYSYS model is converged.
Boundary conditions specifications must be pressure active only.

Figure 16.2: Dynamic specification


Ch 14.hsc - Aspen HYSYS V11 - aspenONE
Flowsheet Case (Main) - Solver Active
Material Stream: 2 Phase Feed
Worksheet
Attachments
Dynamics
Dynamic Specifications
Pressure Specification
Pressure
4.500 bar_g
Active
Flow Specification
Molar
Mass
Std Ideal Liq Vol Flow
Liq Vol Flow @Std Cond
Mass Flow
2.000e+004 kg/h
Feeder block...
OK
Delete
Define from Stream...
View Assay
FCV
Pressure Drop
1.000
bar
PIPE-100
XV-1
FCV
Q-103
2 Phase Feed
Temperature
40.00
C
Pressure
4.500
bar_g
Mass Flow
20000
kg/h
Master Comp Mole Frac (Methane)
0.0200
Master Comp Mole Frac (Ethane)
0.0400
Master Comp Mole Frac (Propane)
0.0400
Master Comp Mole Frac (i-Butane)
0.1000
Master Comp Mole Frac (n-Butane)
0.1000
Master Comp Mole Frac (n-Octane)
0.4000
Master Comp Mole Frac (n-Decane)
0.3000
Pipe length(1)
25.00
m
Elevation(1)
5.000
m
LCV
Pressure
8.000
bar_g
Discharge coefficient
0.6000
Solver (Main) - Ready
View Convergence
Responsiveness: 3
100%

Provide elevation of FCV & XV-01 and make sure that “Pressure Flow Relation” is checked. As per line sizing done in steady-state, provide the same pipe diameters in the “Rating\Nozzles\Nozzle Parameters” tab.

Figure 16.3: FCV

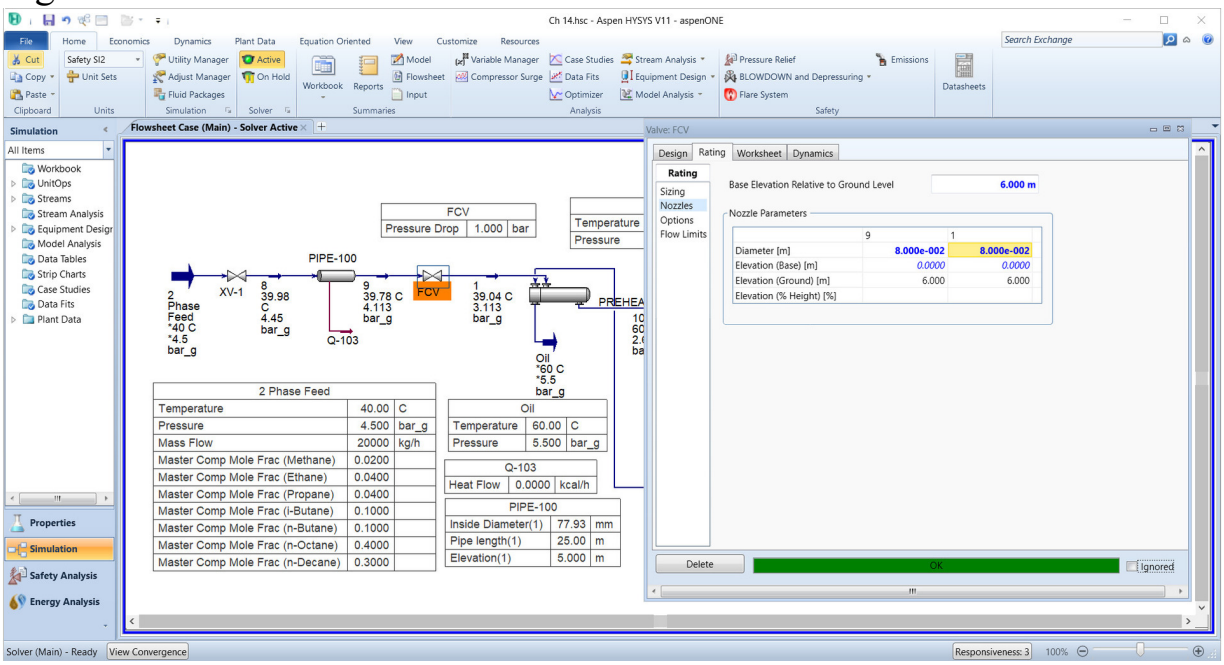


Figure 16.4: FCV

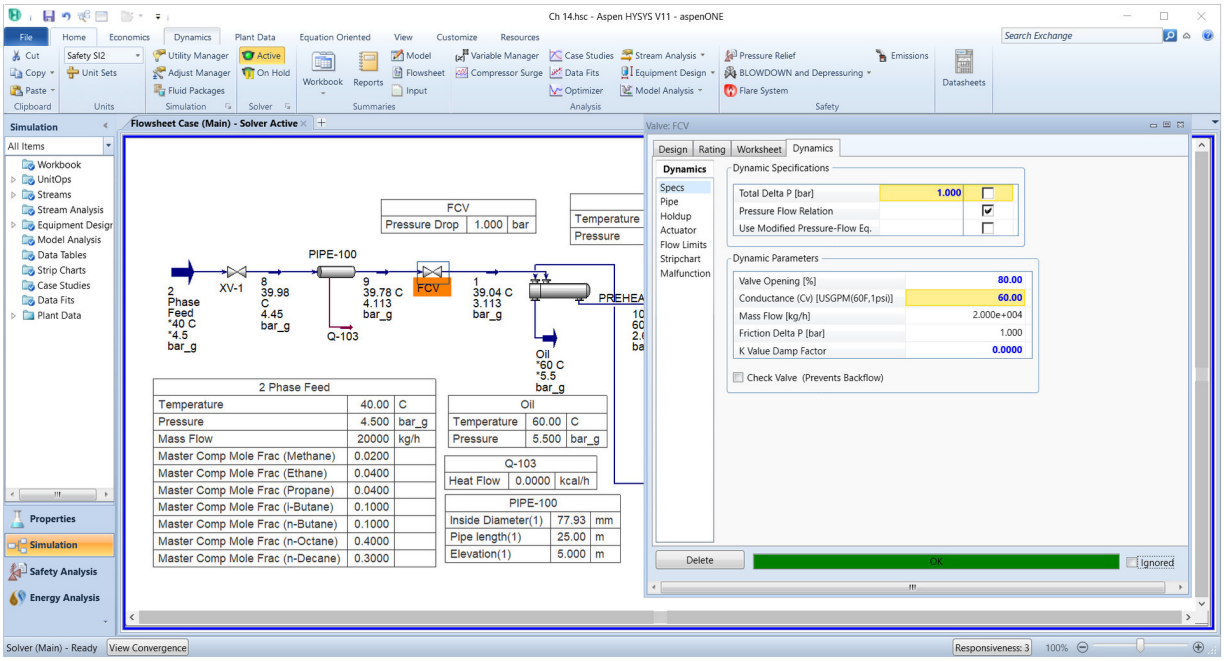
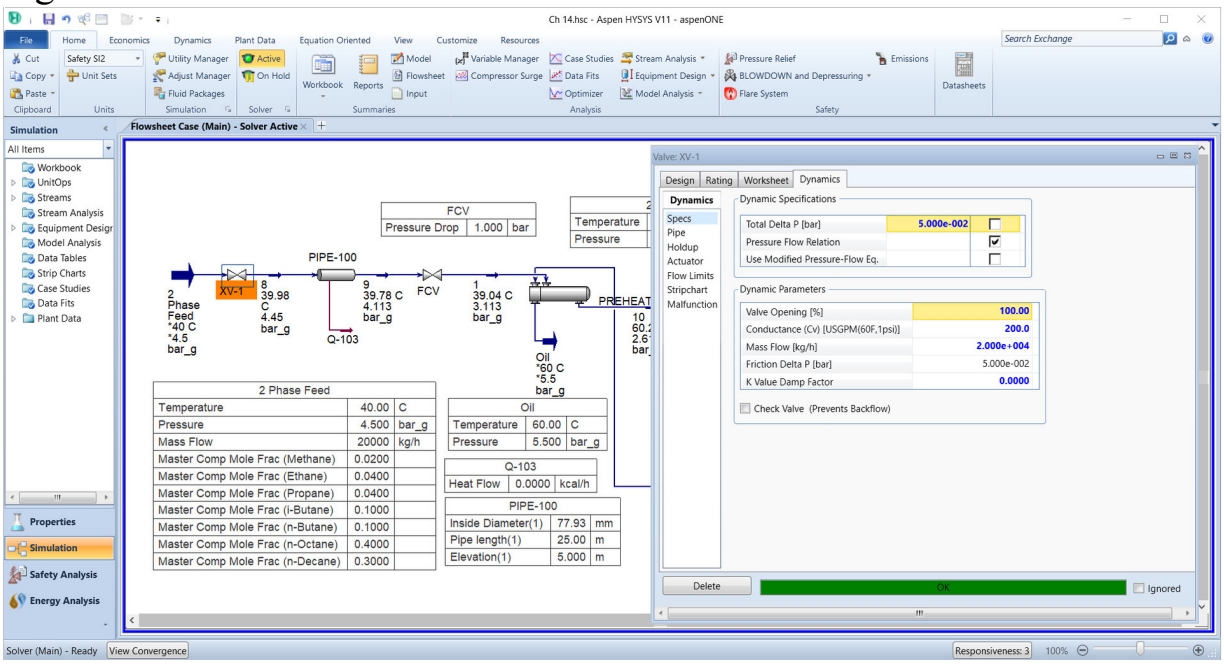
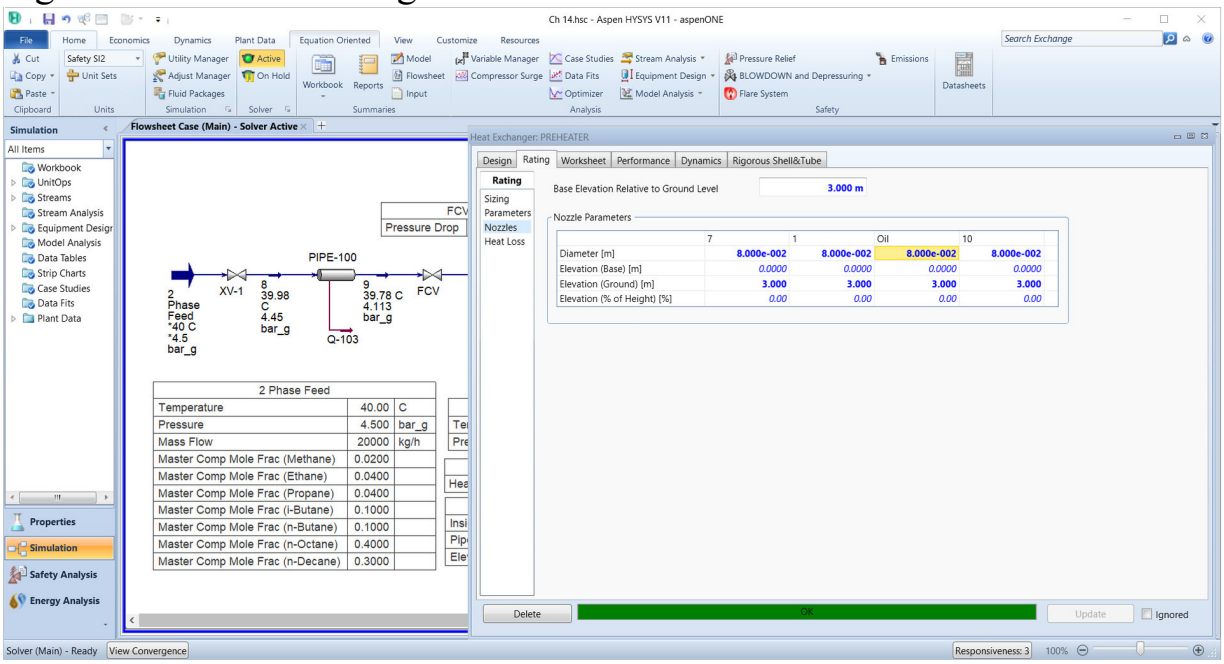


Figure 16.5: XV-1



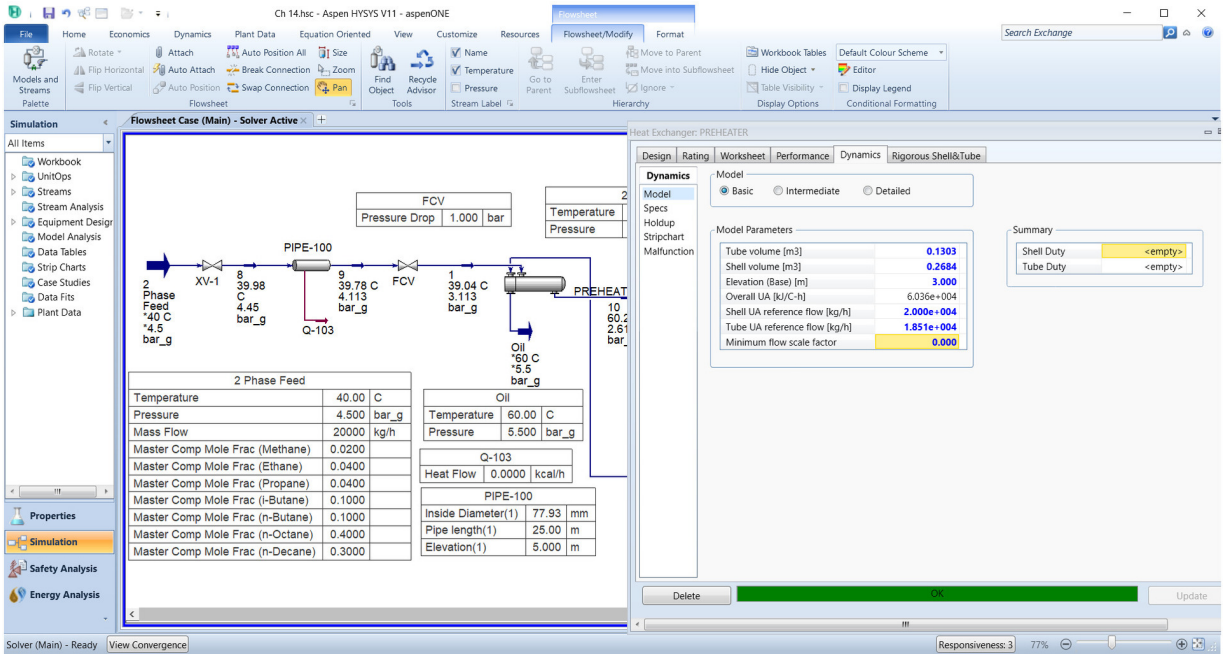
Provide elevation and pipe diameters for the pre-heater (E-100).

Figure 16.6: Heat exchanger



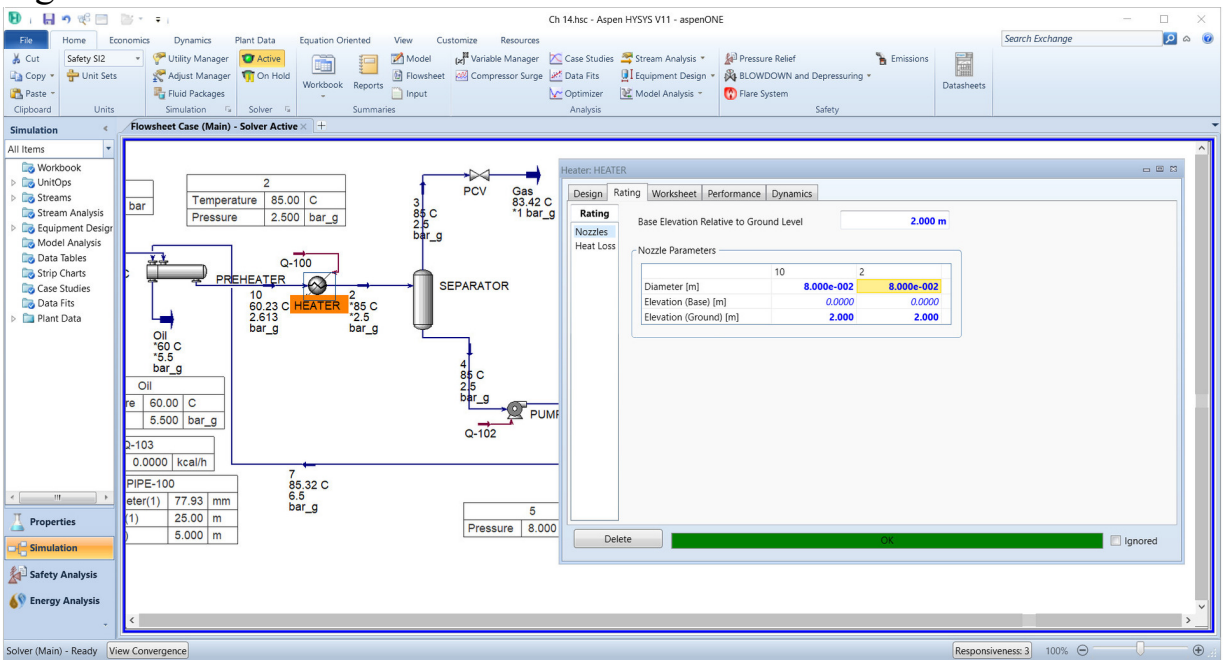
Tube volume and shell volume are provided (it will be available in EDR output) along with other details as given below.

Figure 16.7: Heat exchanger



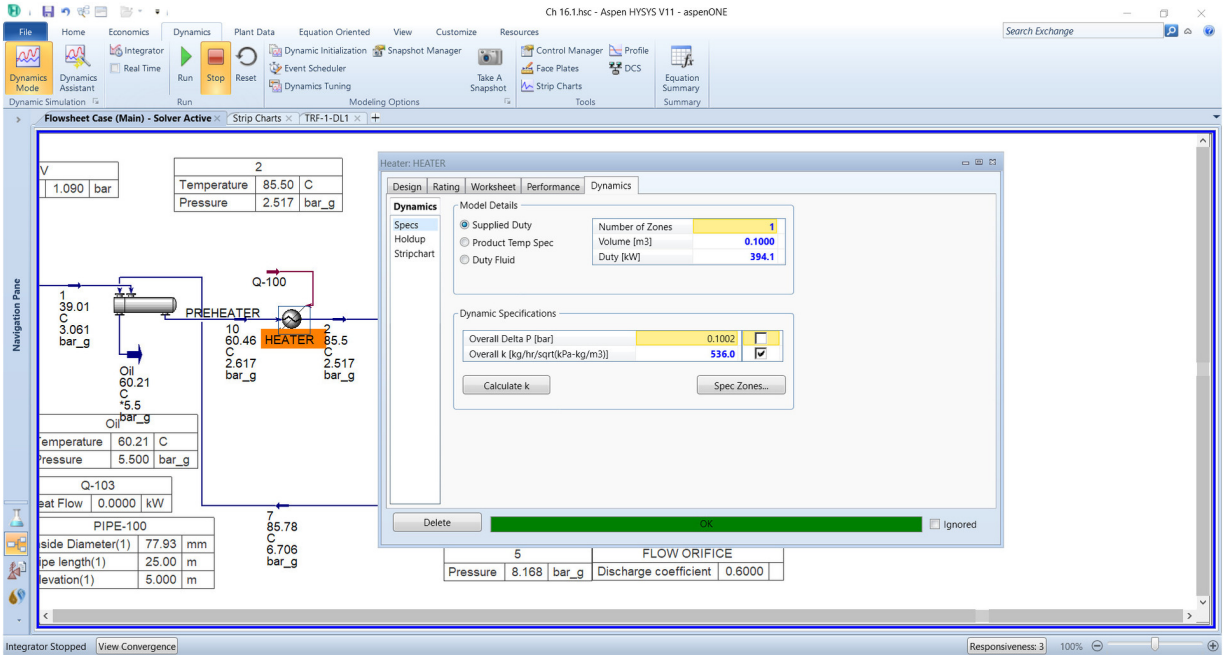
Provide elevation and pipe diameters for heater (E-100).

Figure 16.8: Heater



For assigned pressure drop value, “Overall k” will be calculated by HYSYS after clicking “Calculate k”. This enables HYSYS to calculate pressure drop at different flow rates through the heater using the “Overall k” value.

Figure 16.9: Heater



Provide separator dimensions and elevation.

Figure 16.10: Separator

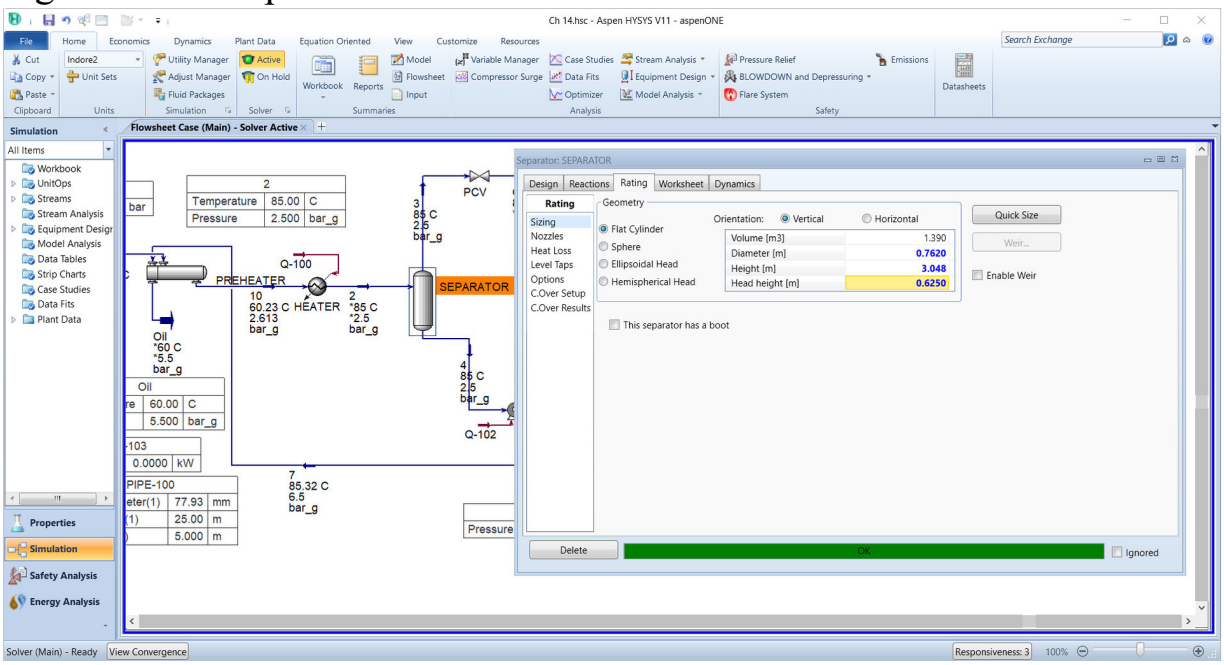
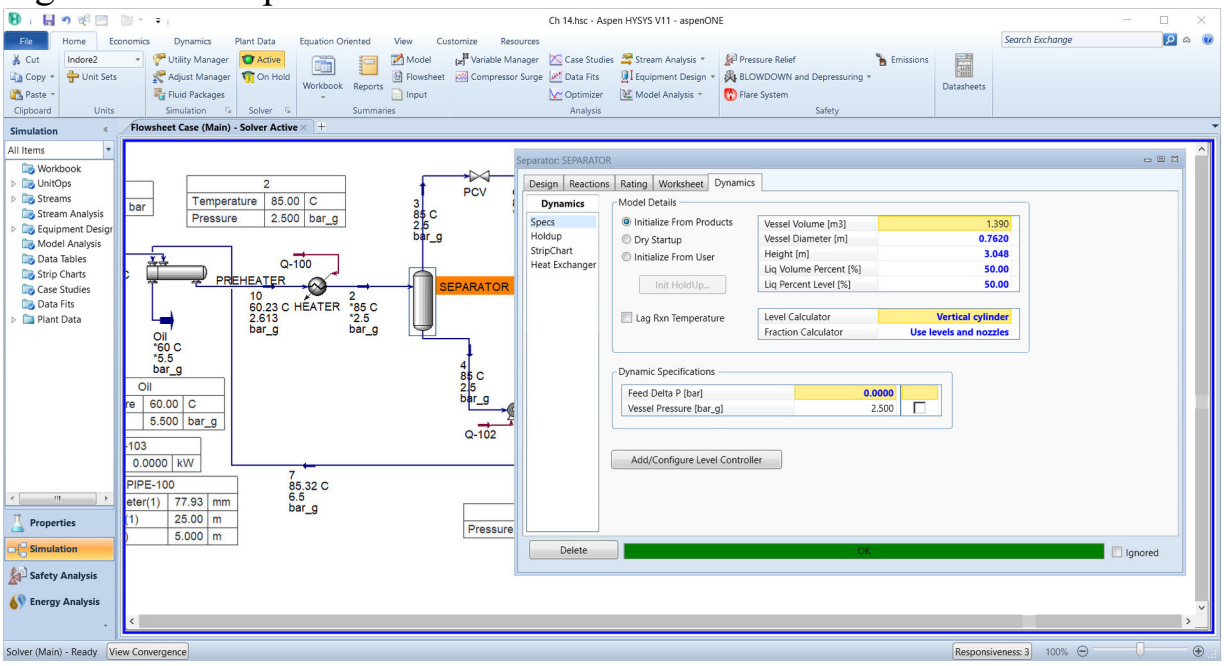


Figure 16.11: Separator



Provide PCV elevation and pipe in/out diameter.

Figure 16.12: PCV

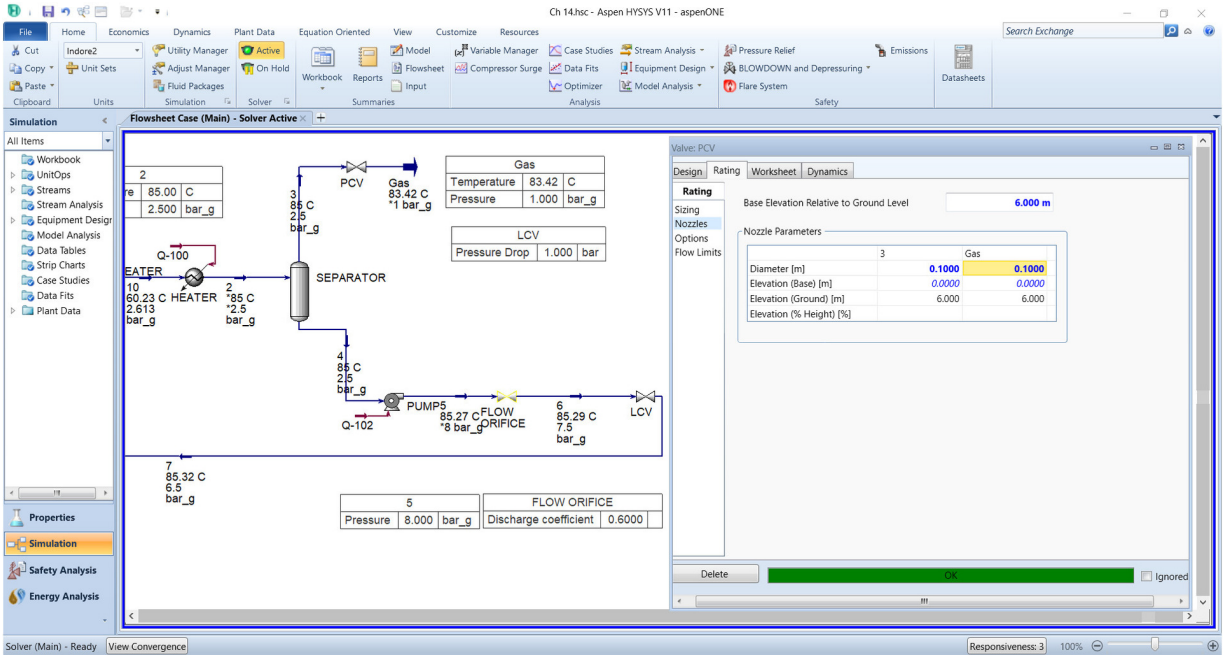
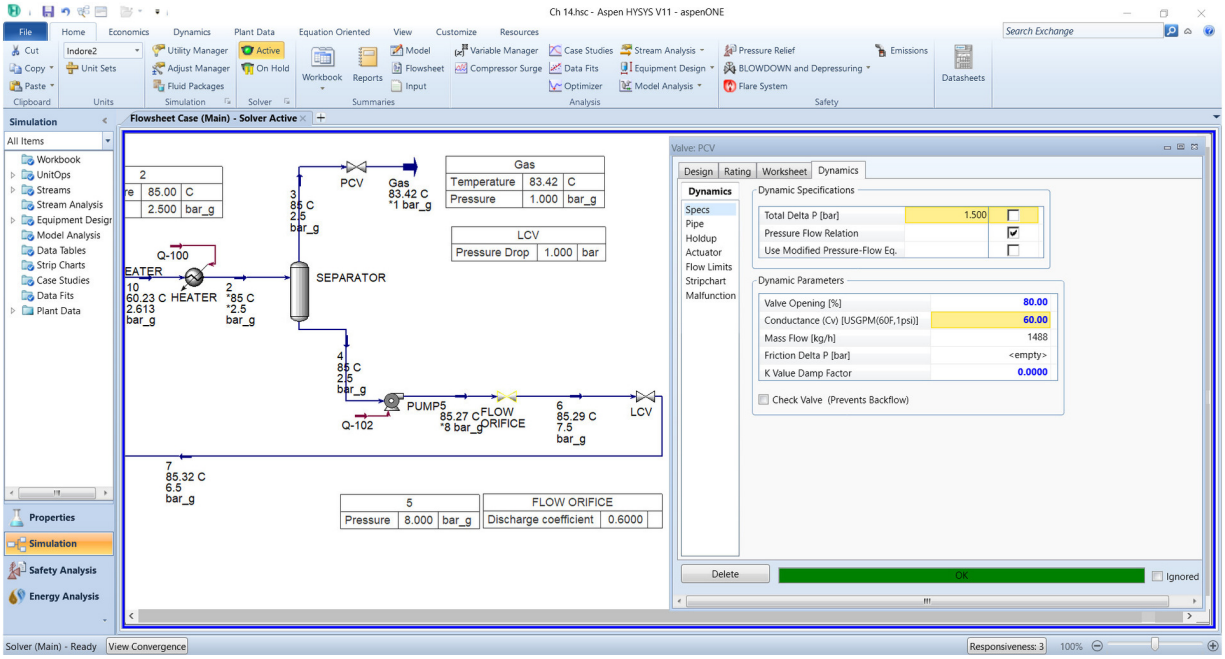


Figure 16.13: PCV



Provide pump elevation, in/out pipe diameter and pump curve data.

Figure 16.14: Pump

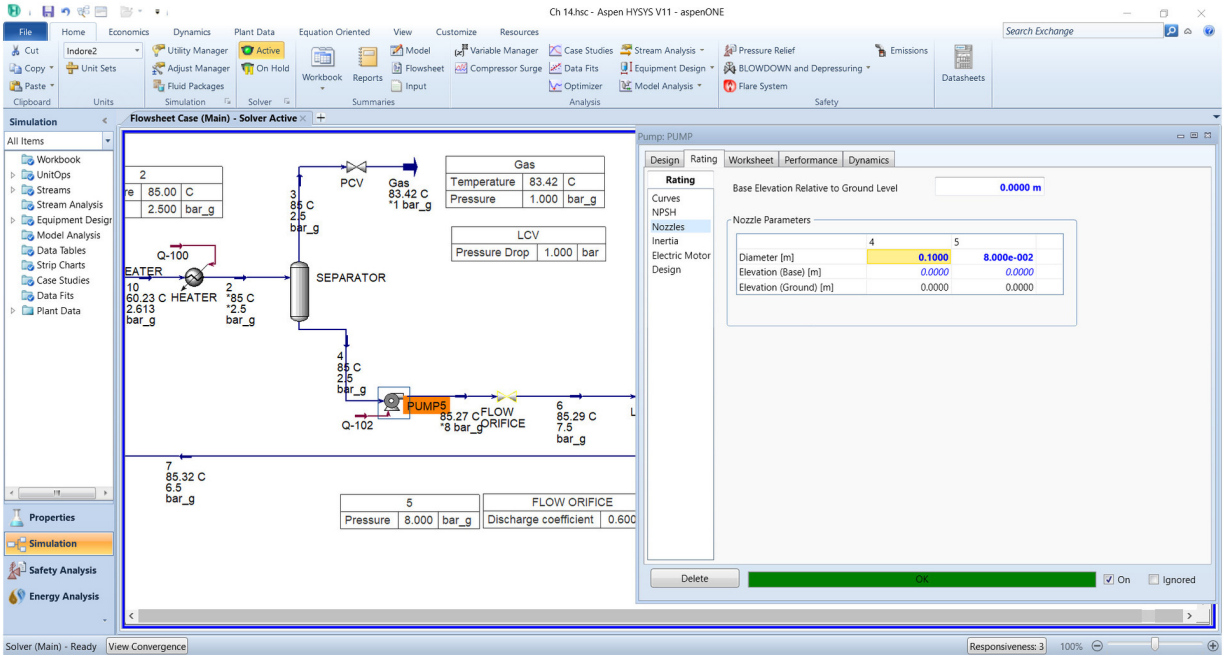


Figure 16.15: Pump

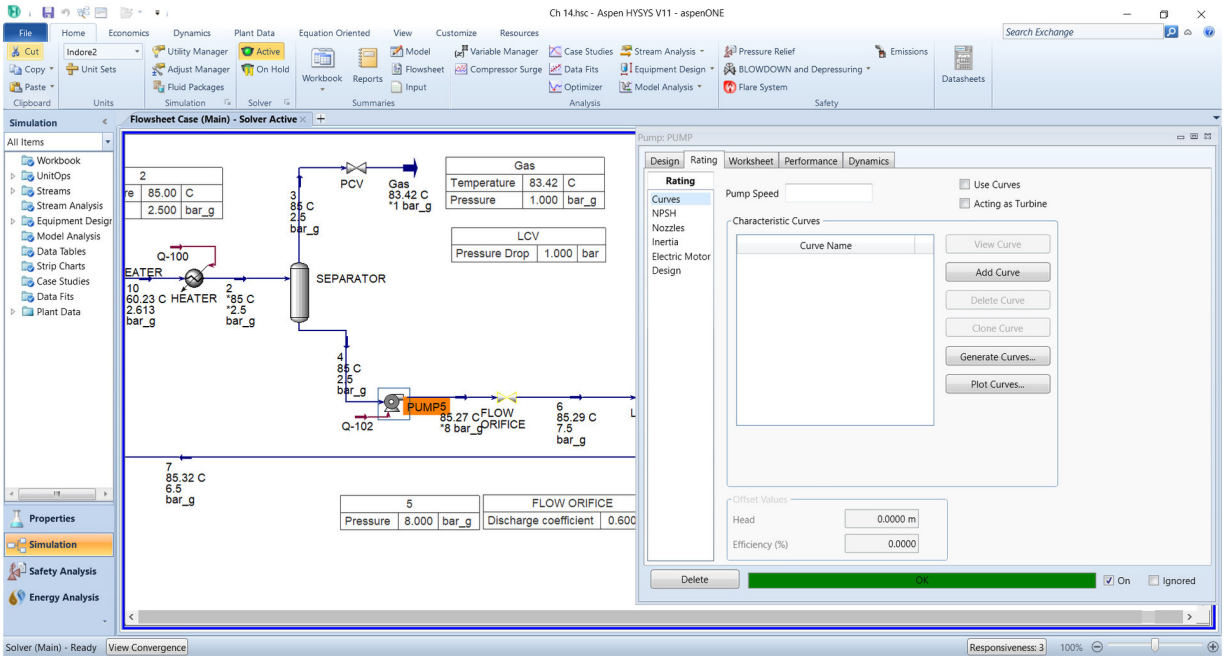


Figure 16.16: Pump

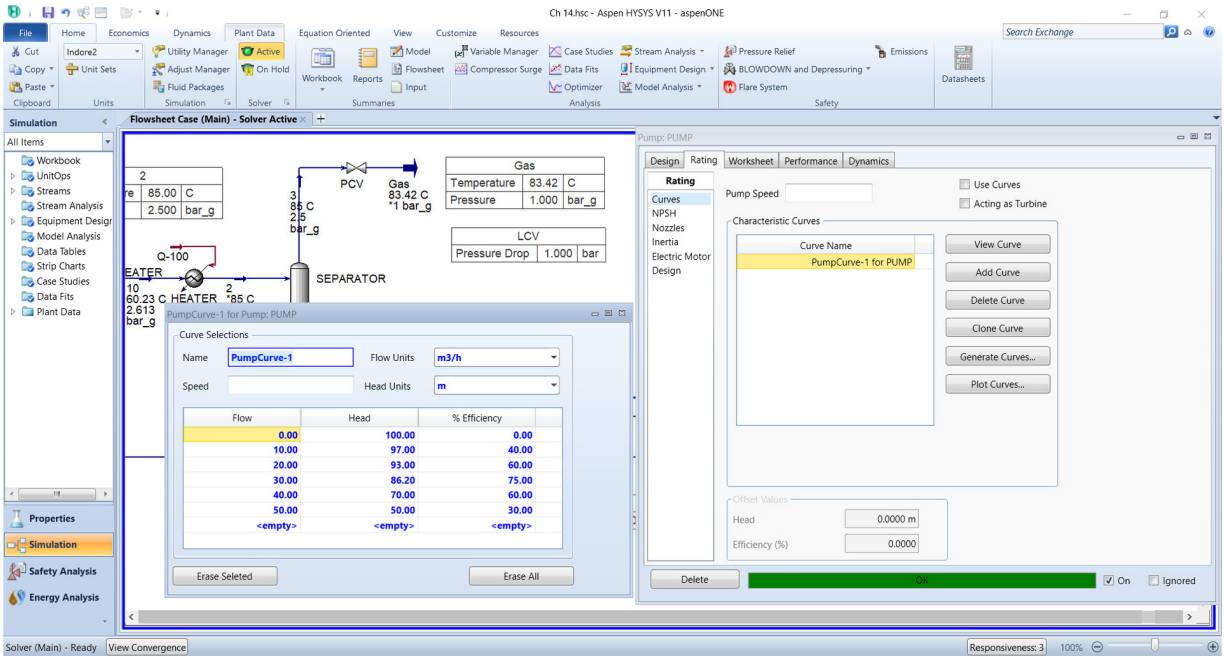


Figure 16.17: Pump

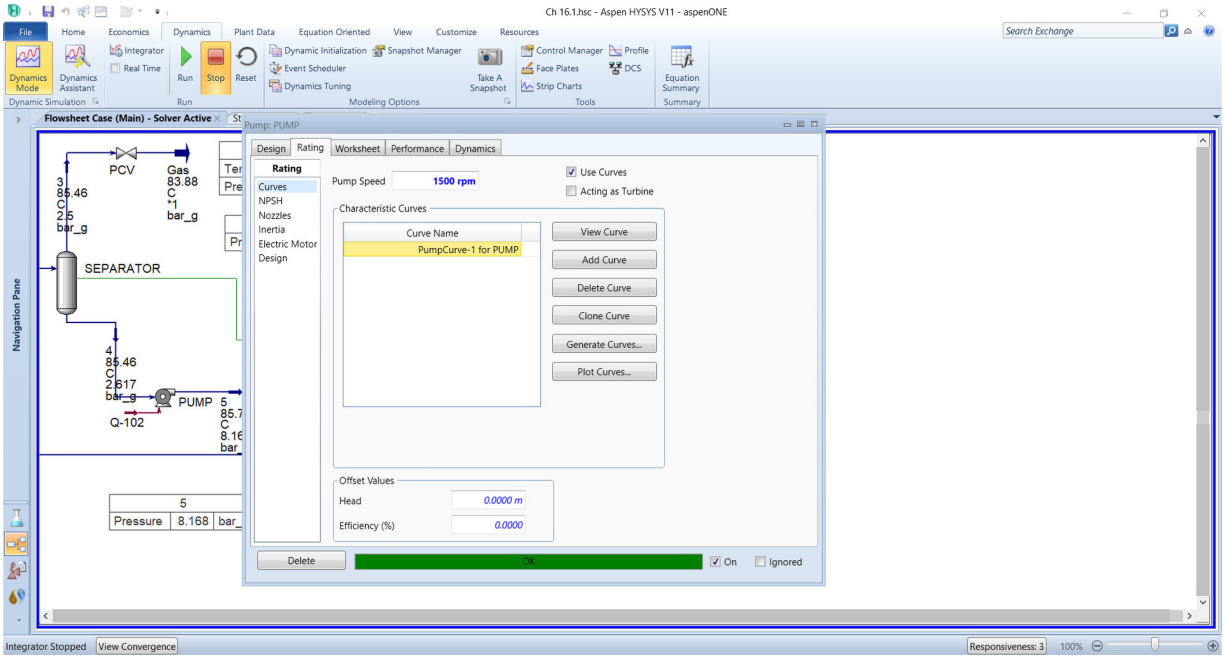
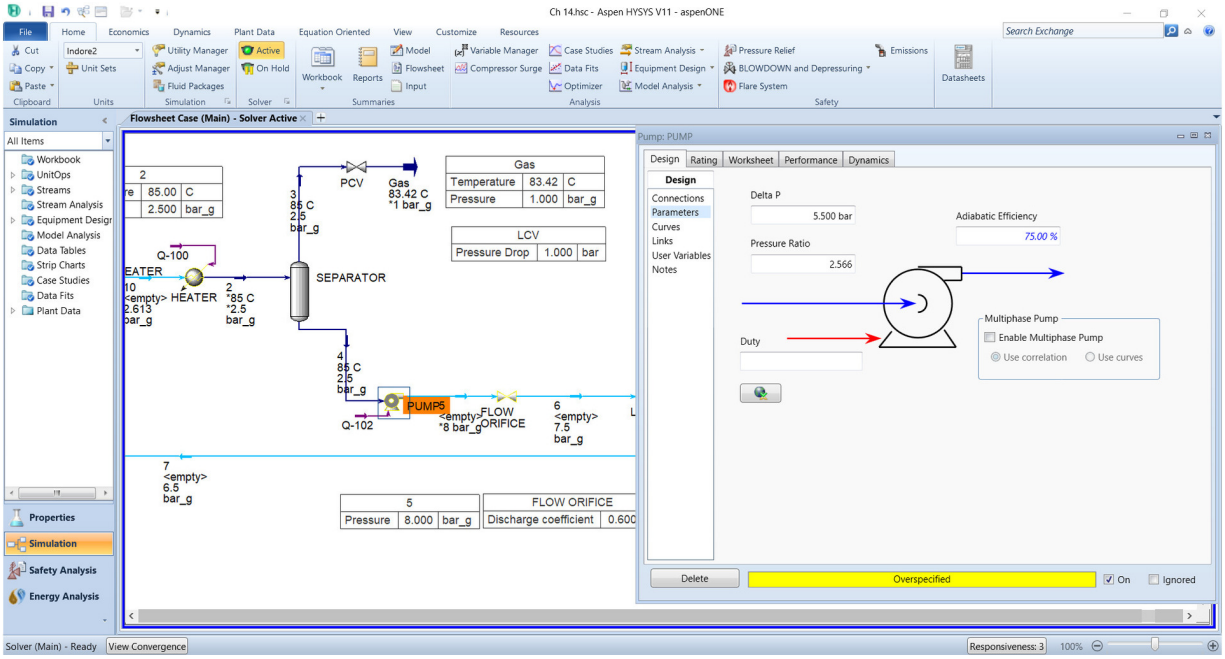


Figure 16.18: Pump



Delete the pump efficiency and pump discharge pressure.

Figure 16.19: Pump

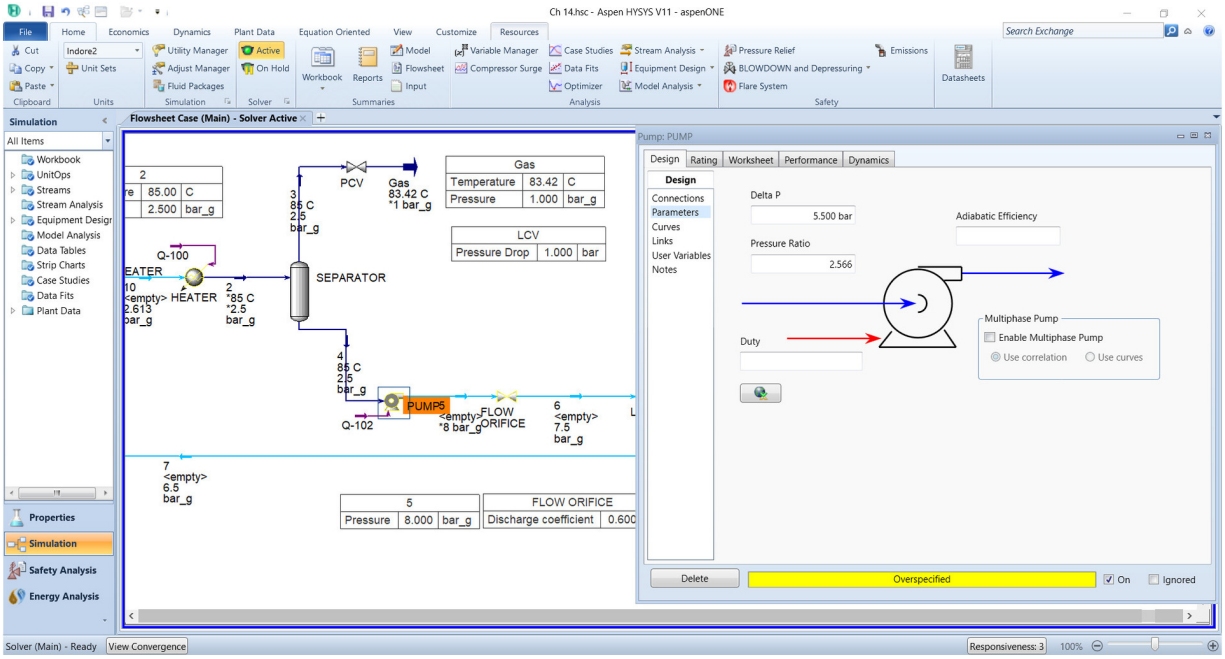


Figure 16.20: Pump

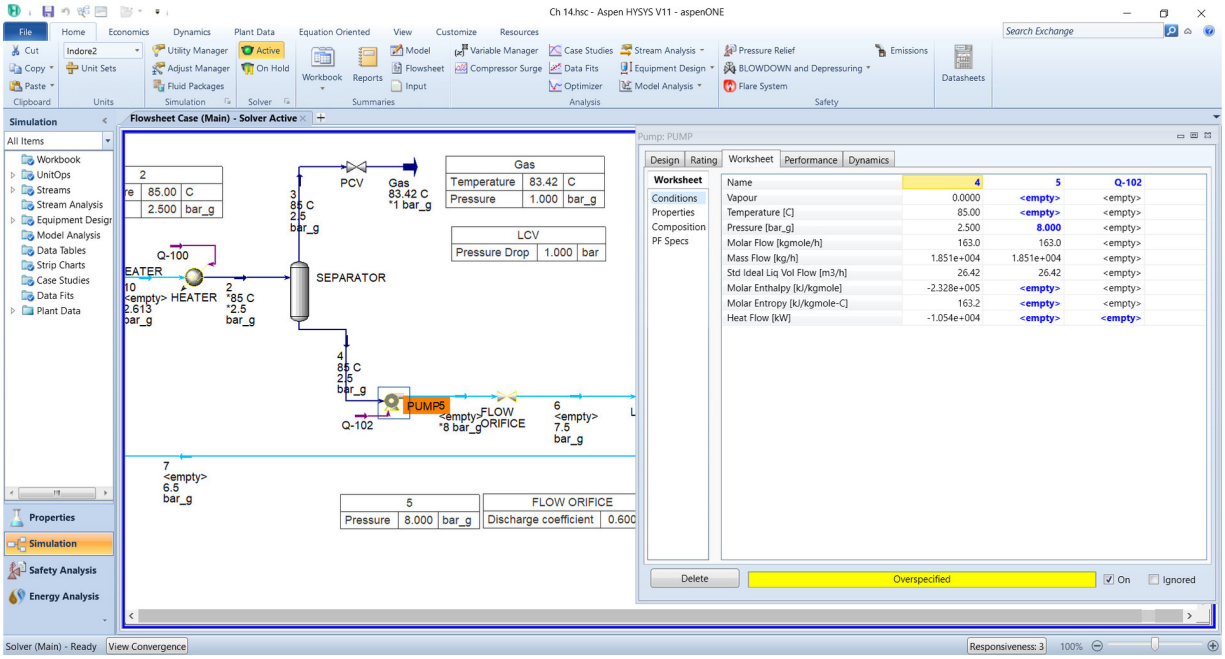


Figure 16.21: Pump

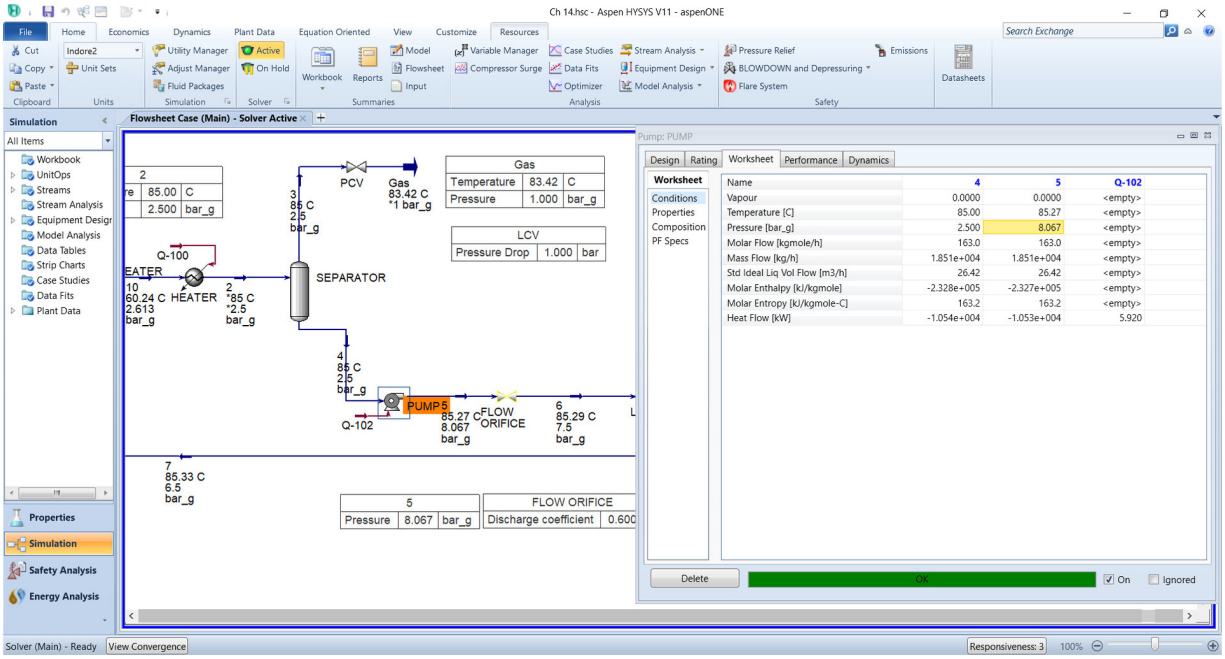
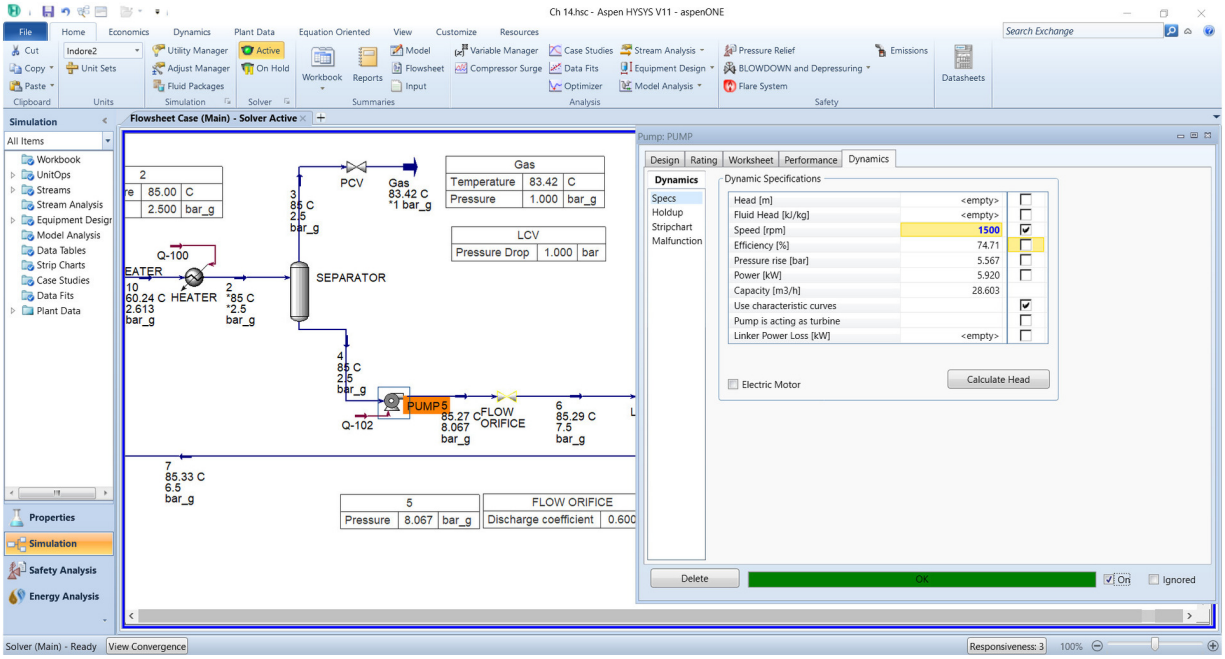


Figure 16.22: Pump



Provide flow orifice elevation and set pressure & full open pressure value as 0 barg.

Figure 16.23: Orifice

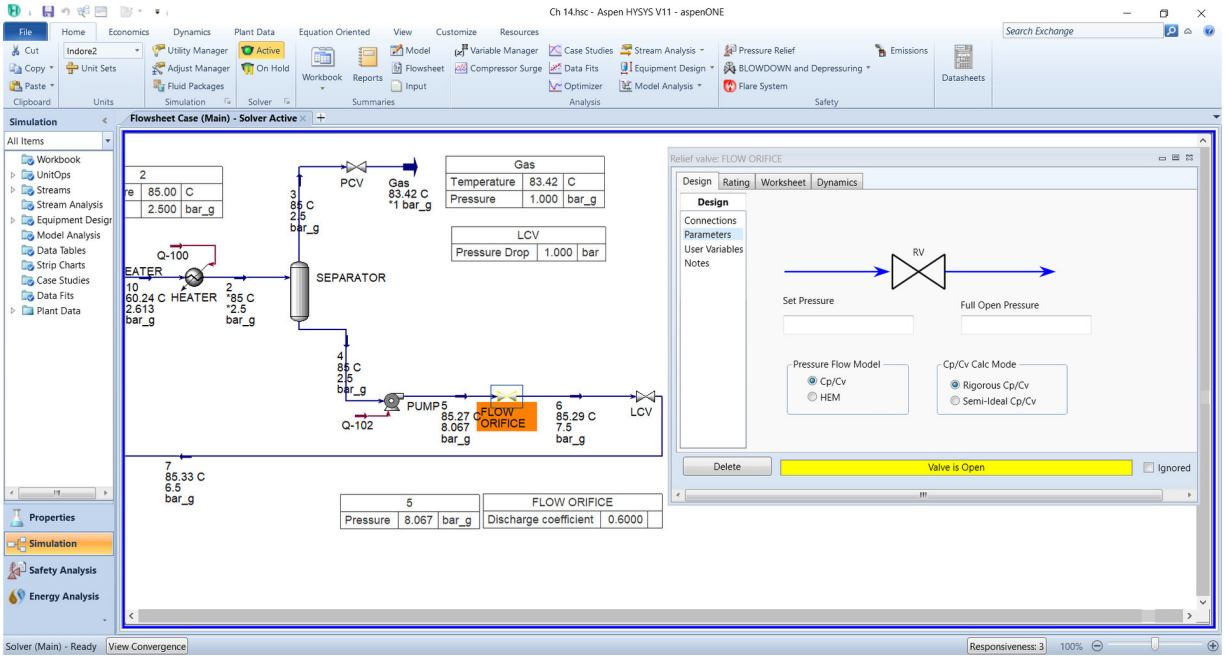
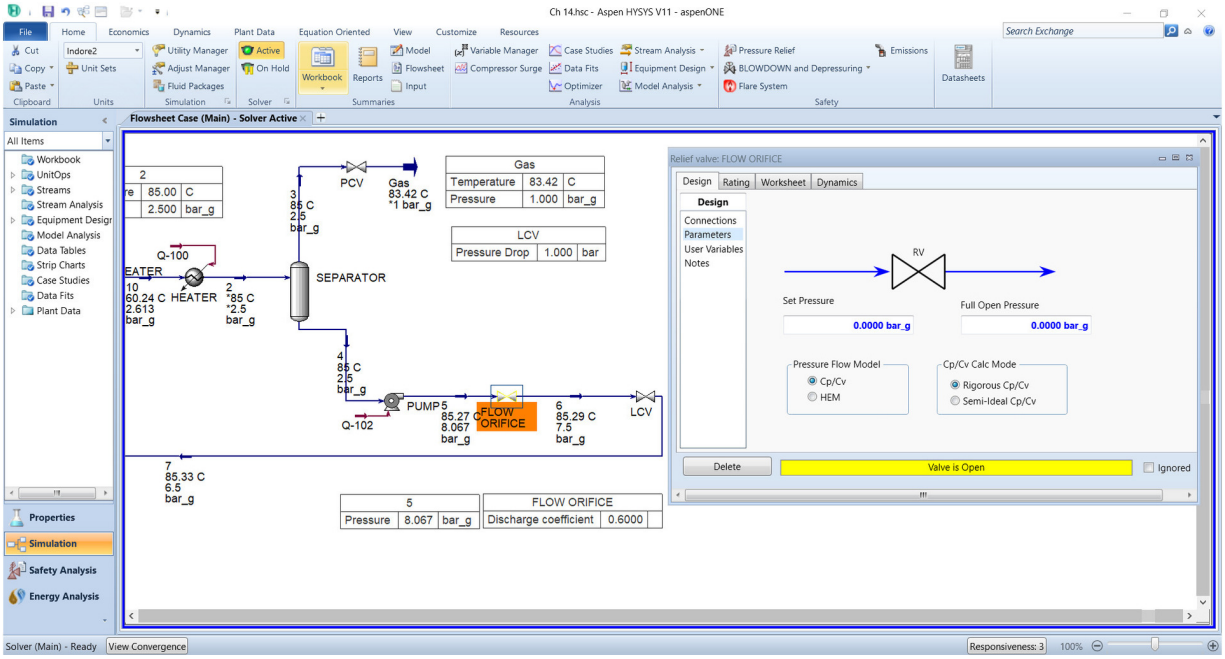


Figure 16.24: Orifice



Provide LCV elevation and pipe in/outsize.

Figure 16.25: LCV

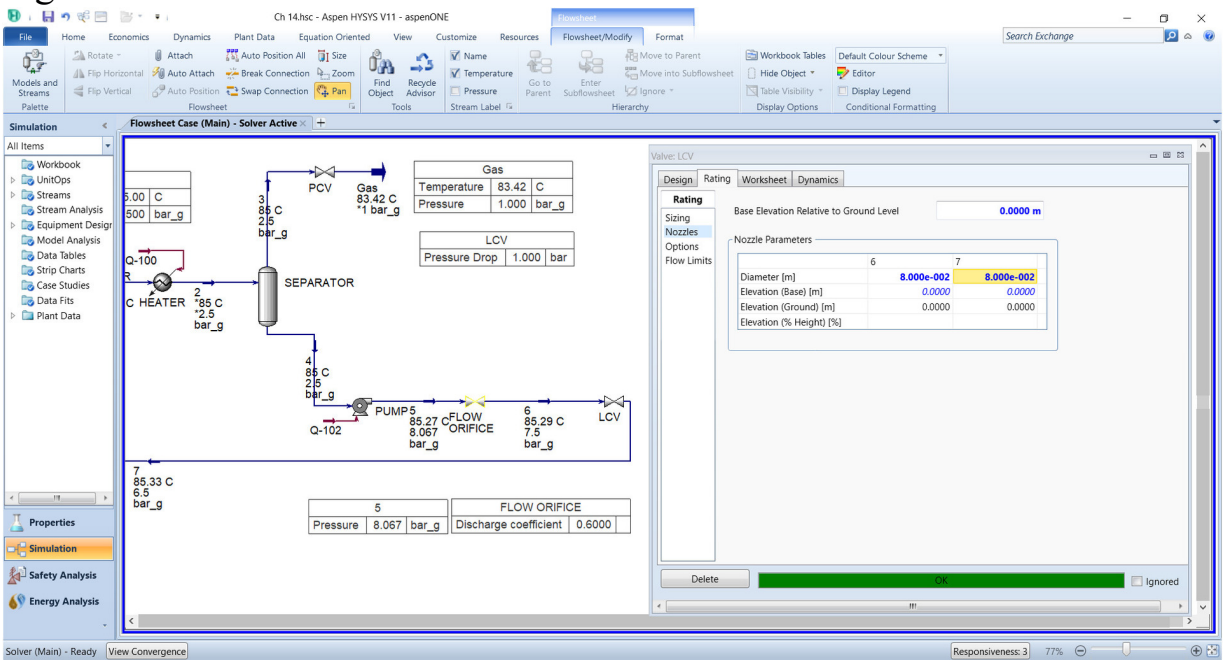
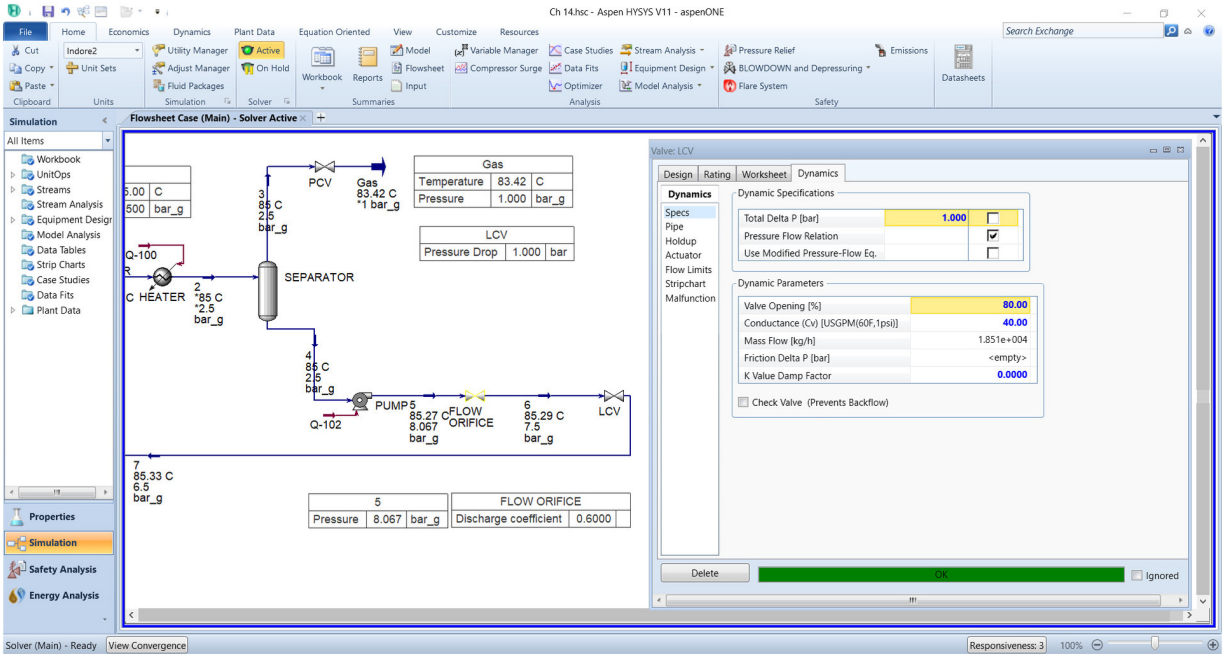


Figure 16.26: LCV



Provide FIC-1, TIC-1, LIC-1, and PIC-1.

Figure 16.27: Controllers

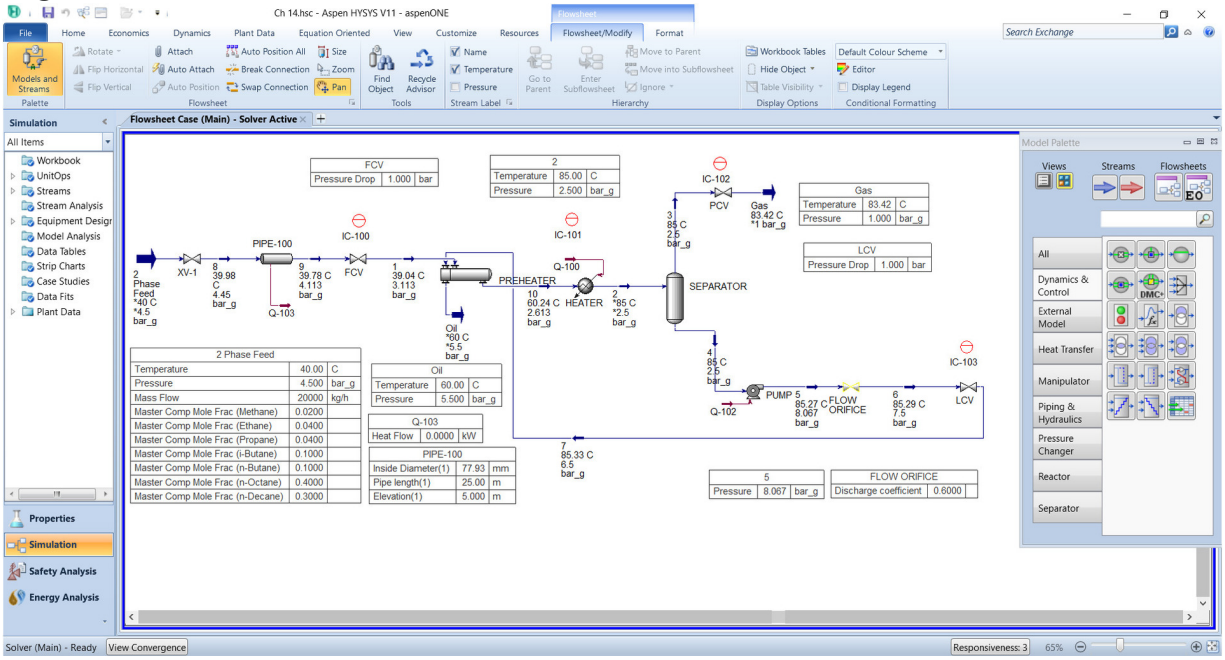
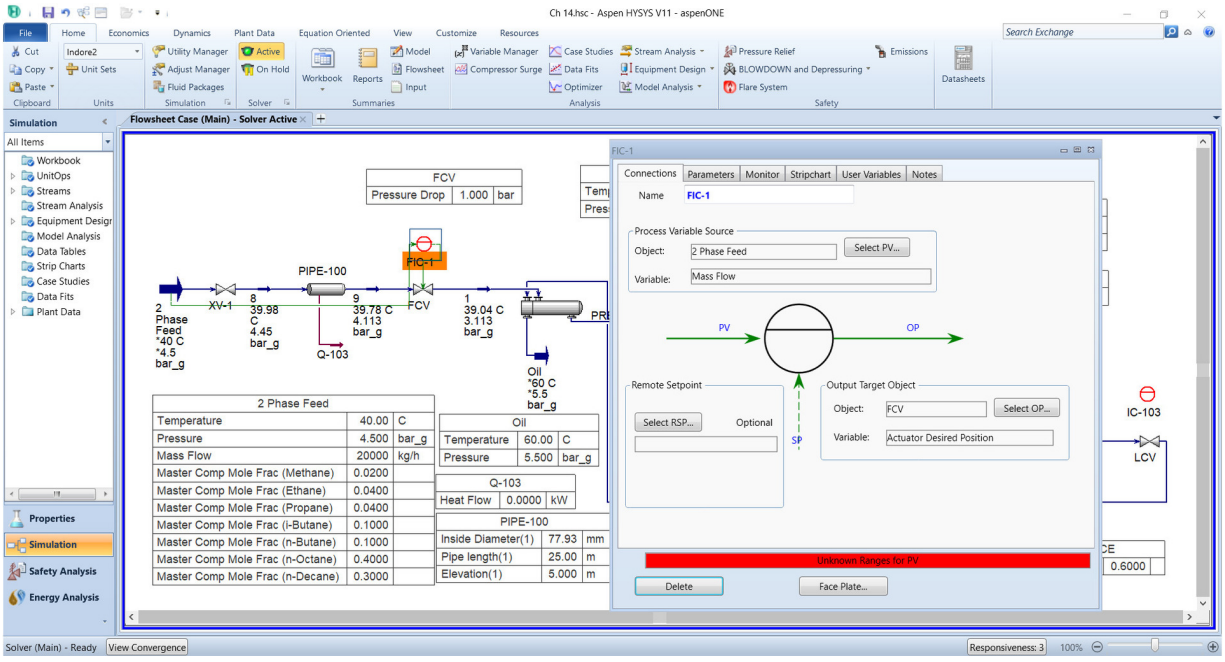


Figure 16.28: FIC



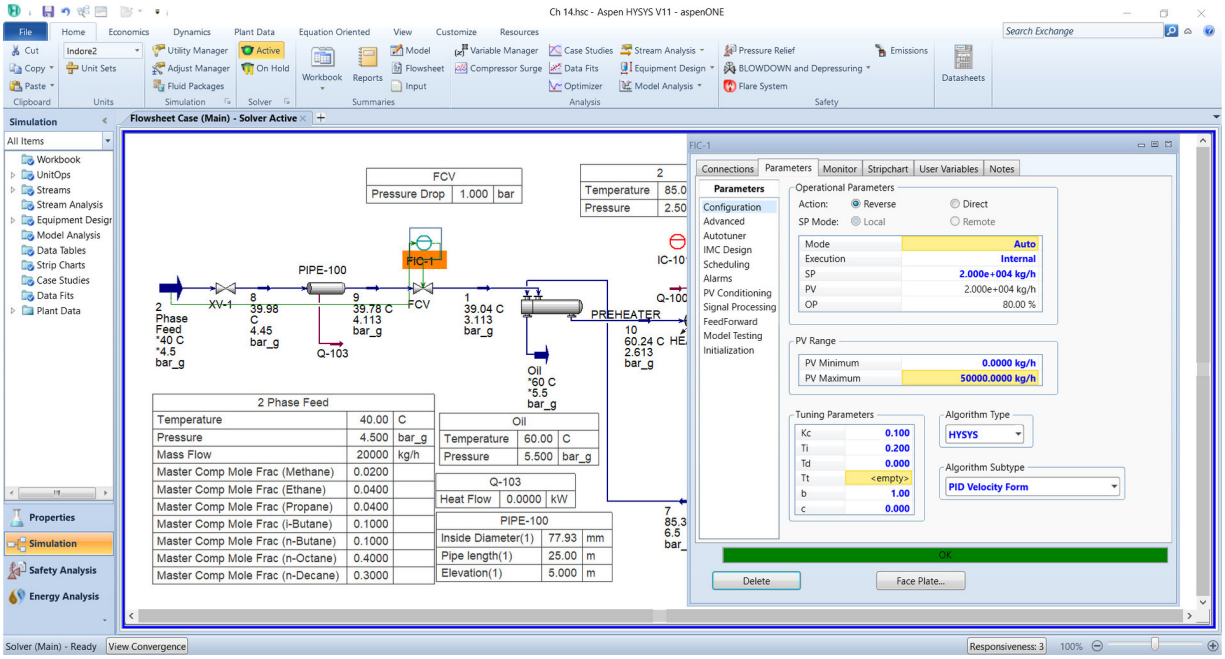
There is a set of preliminary tuning parameters suggested. The same shall be used in this example.

Figure 16.29: Tuning parameters

System	K_c	τ_i (minutes)	τ_d (minutes)
Flow	0.1	0.2	0
Level	2	10	0
Pressure	2	2	0
Temperature	1	20	0

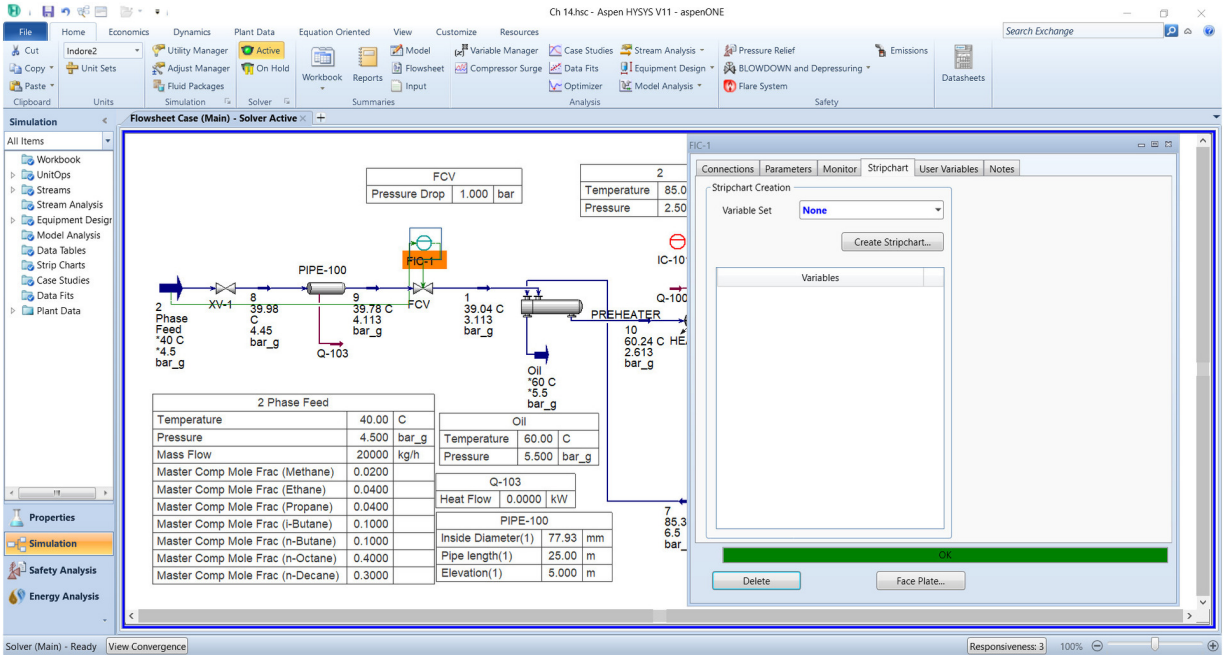
FIC-1 details are provided as below;

Figure 16.30: FIC



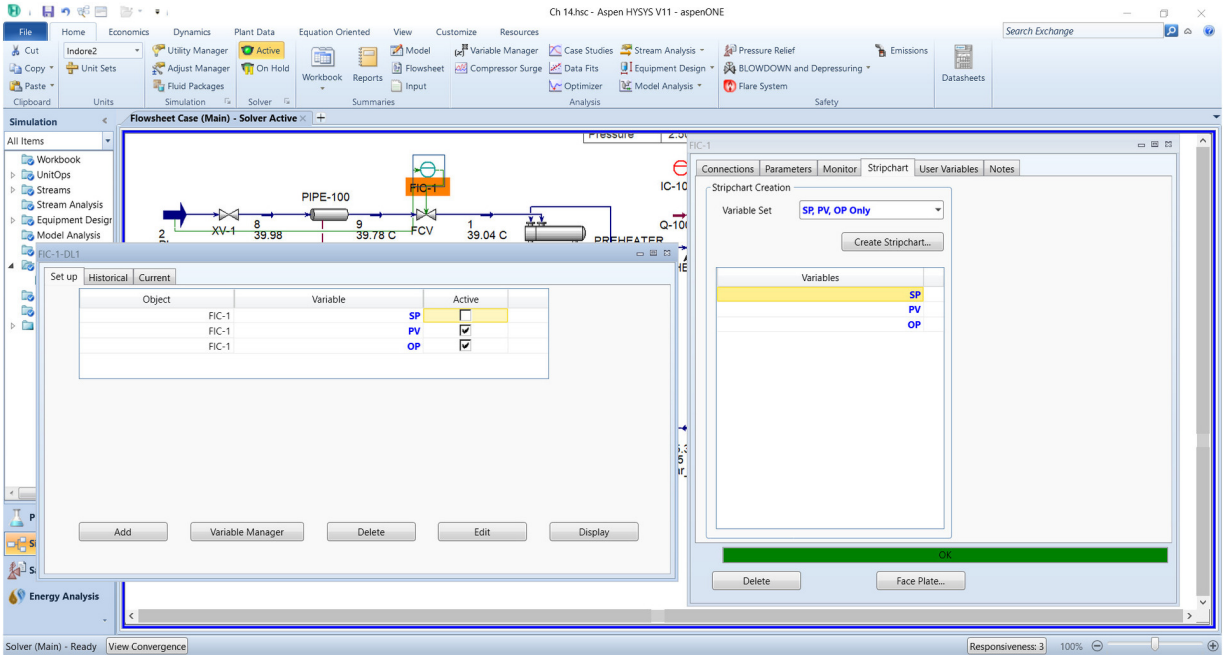
Strip charts are mandatory in dynamic simulation to analyze the model performance.
Select “Strip chart” and “Create Strip chart”.

Figure 16.31: Strip chart



Uncheck the “SP” box and click on the “Display” tab.

Figure 16.32: Strip chart



Right-click on the plot and select “Graph Control”.

Figure 16.33: Strip chart

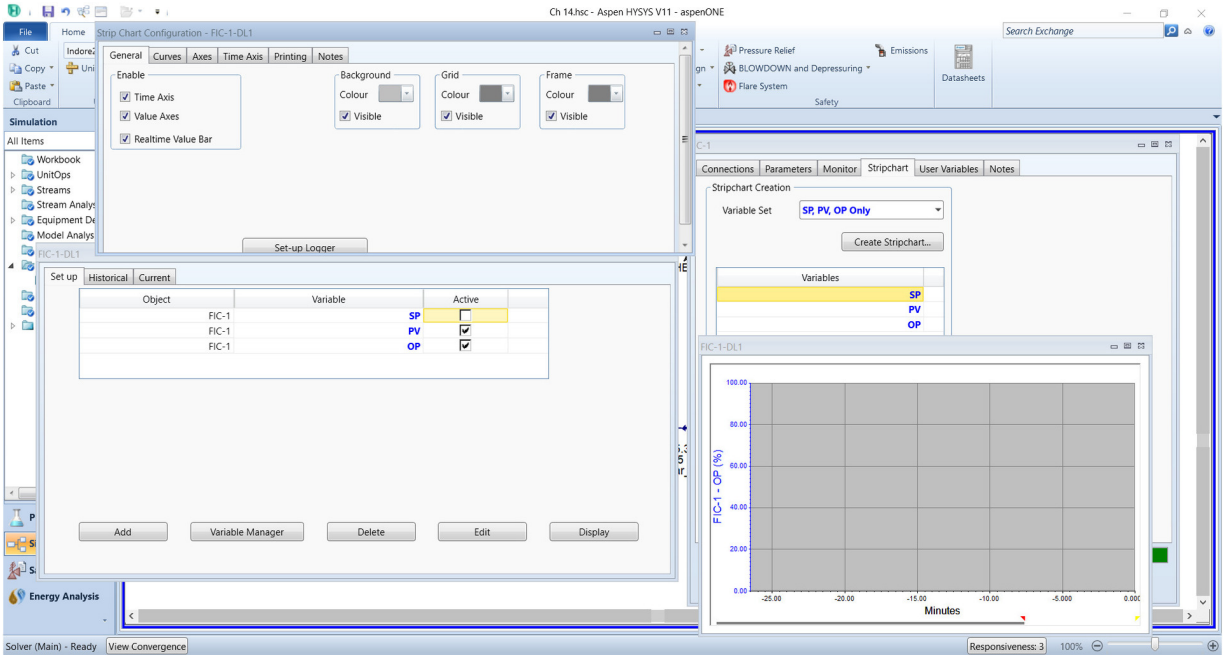


Figure 16.34: Strip chart

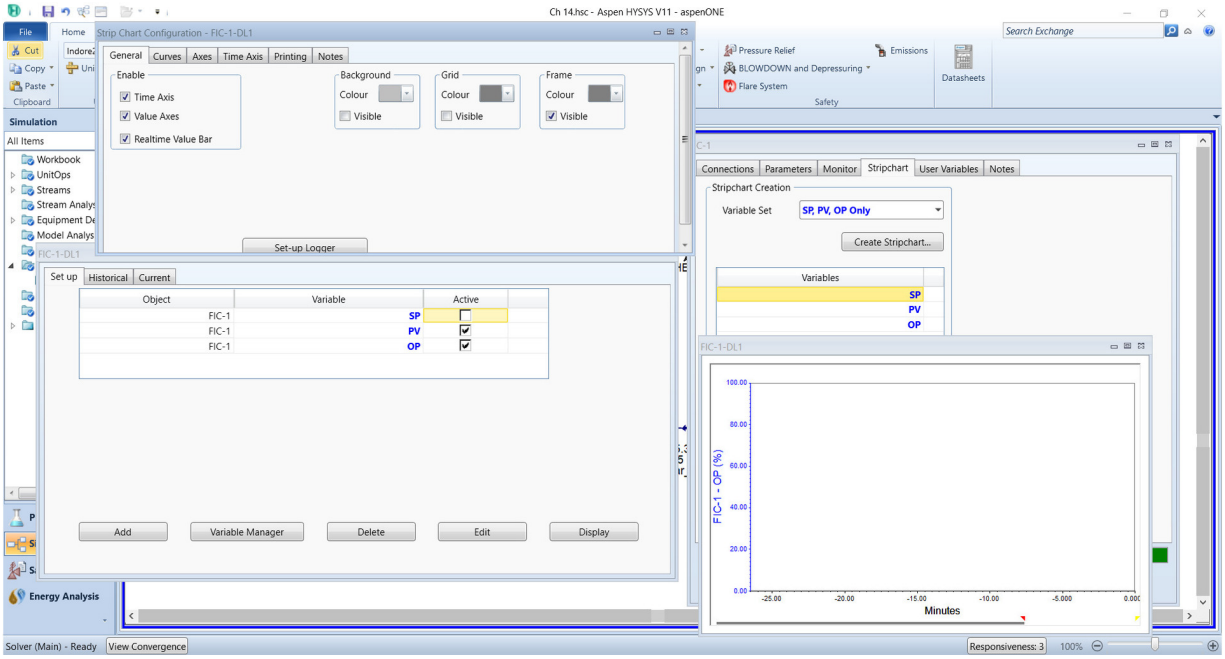


Figure 16.35: Strip chart

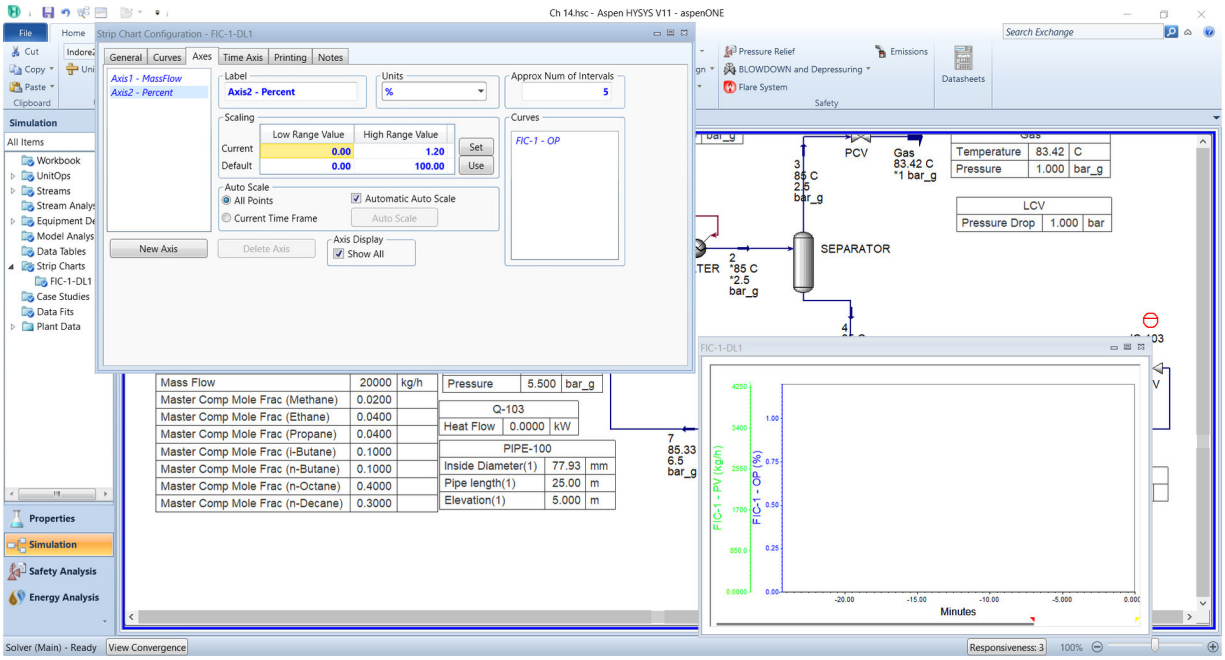


Figure 16.36: TIC

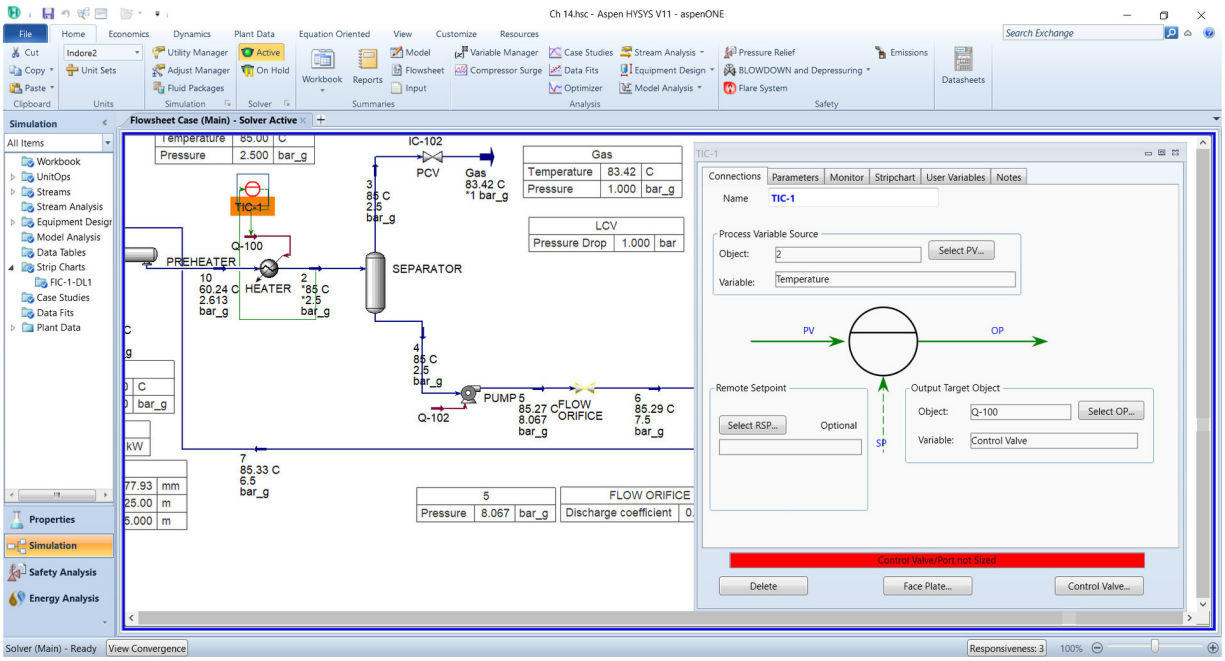


Figure 16.37: TIC

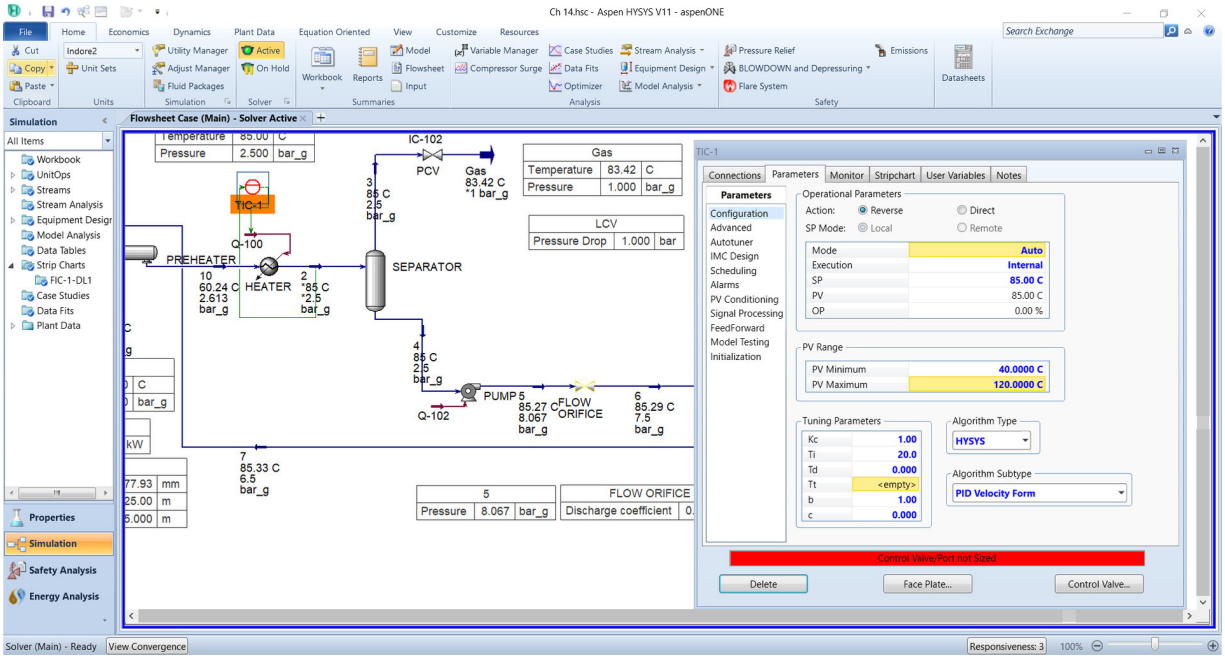


Figure 16.38: TIC

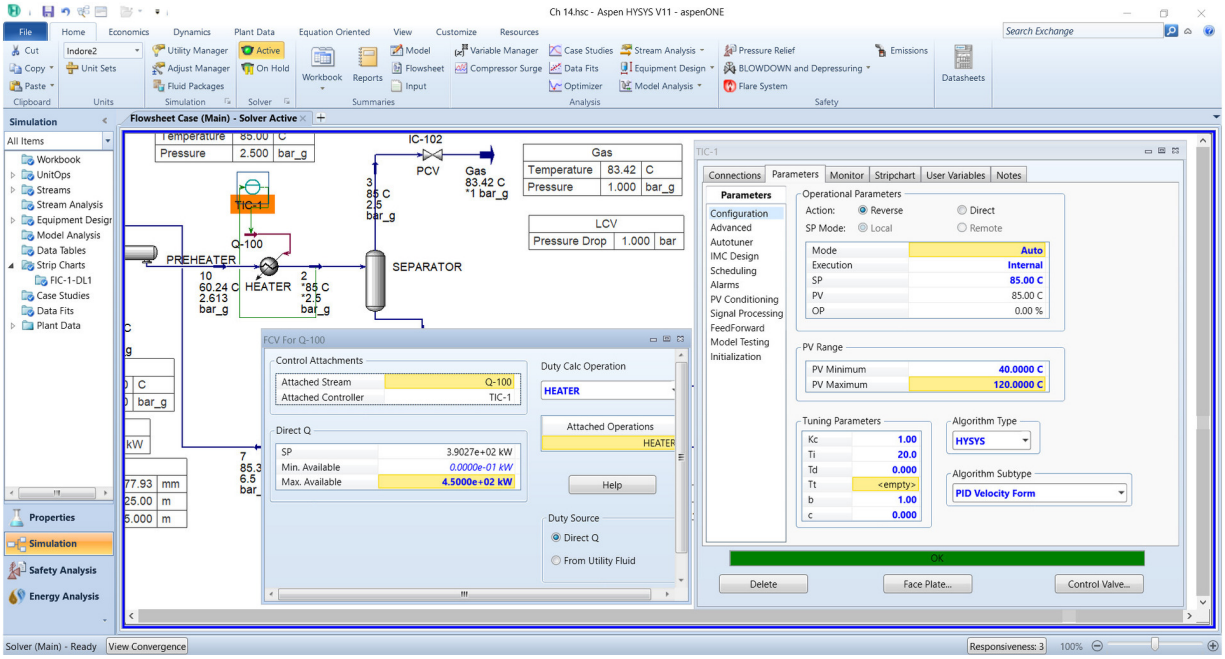


Figure 16.39: TIC

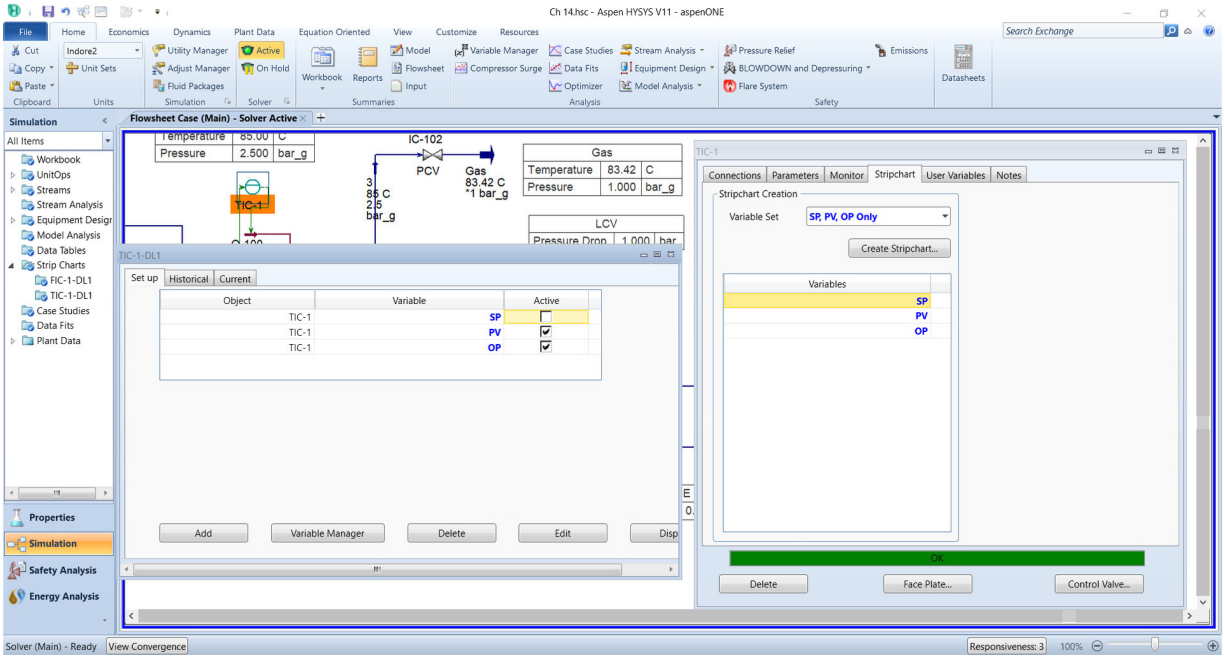


Figure 16.40: PIC

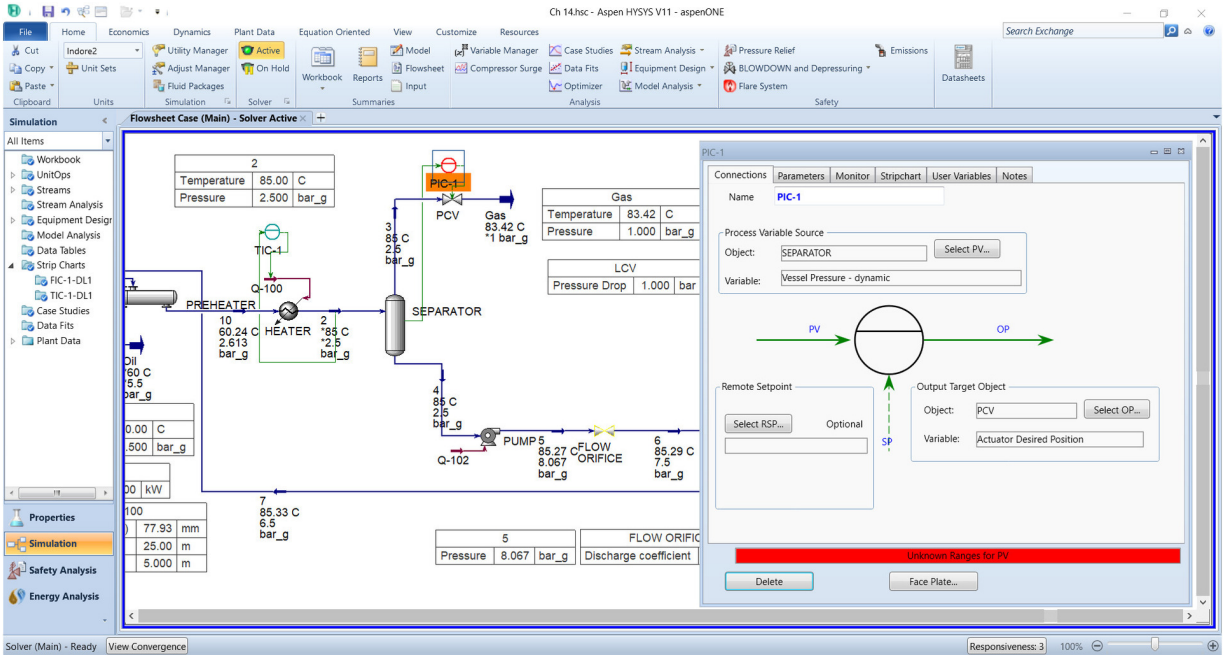


Figure 16.41: PIC

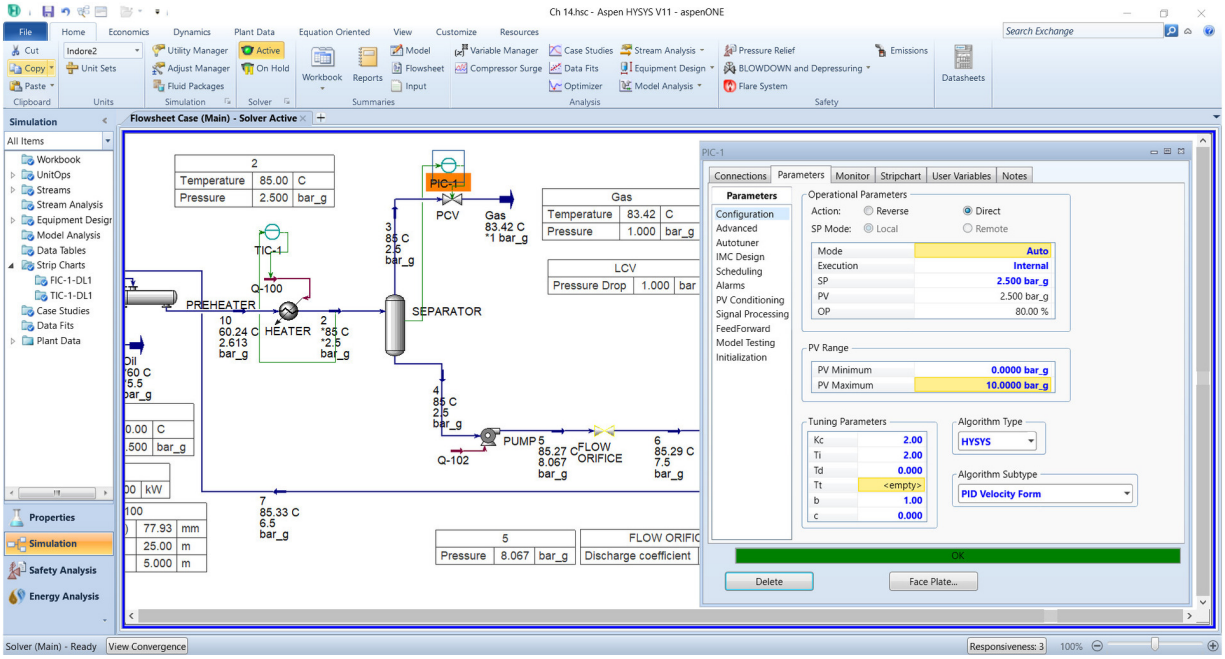


Figure 16.42: PIC

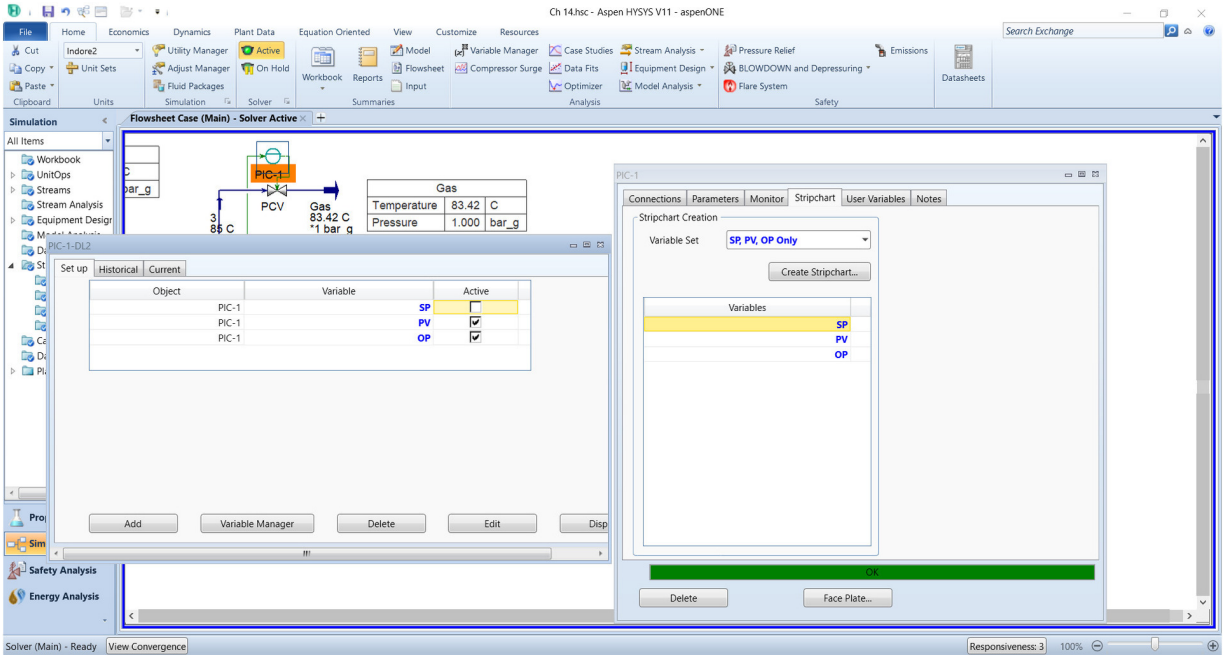


Figure 16.43: LIC

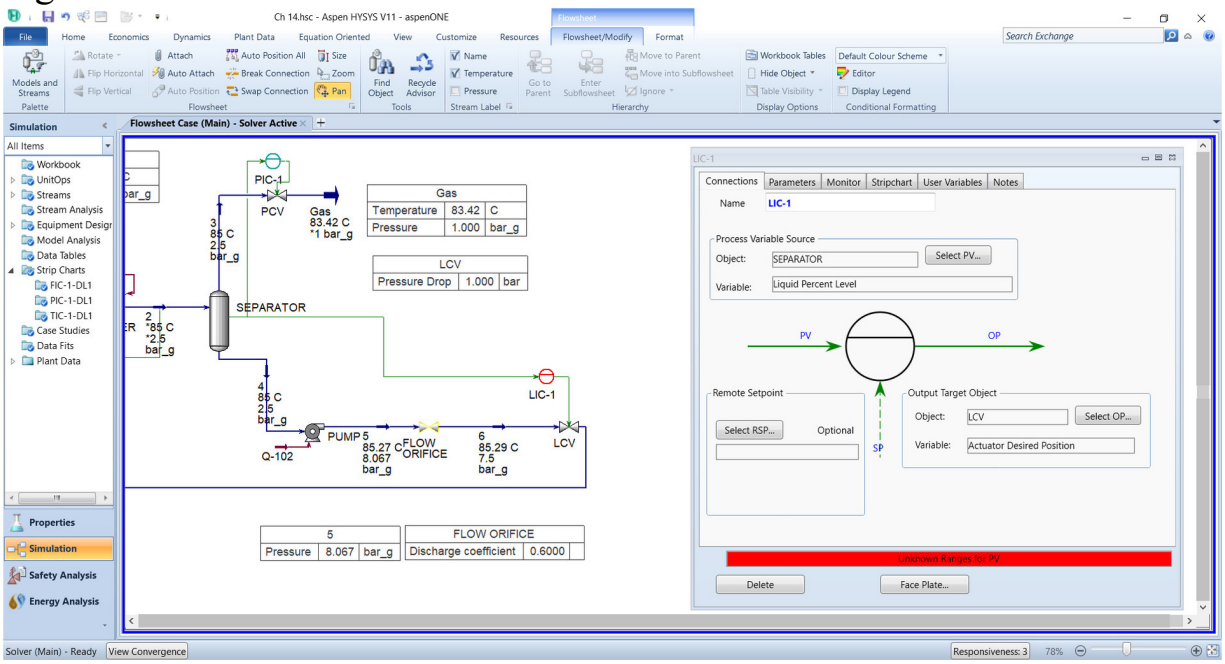
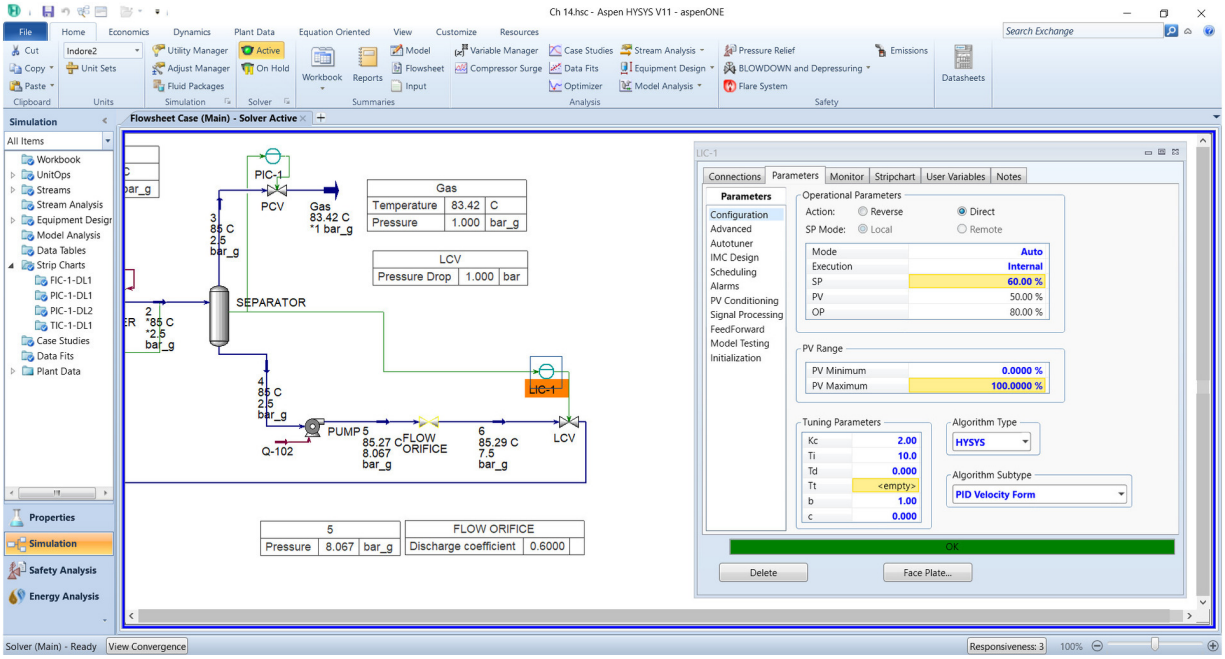


Figure 16.44: LIC



Provide the strip charts.

Figure 16.45: Strip chart

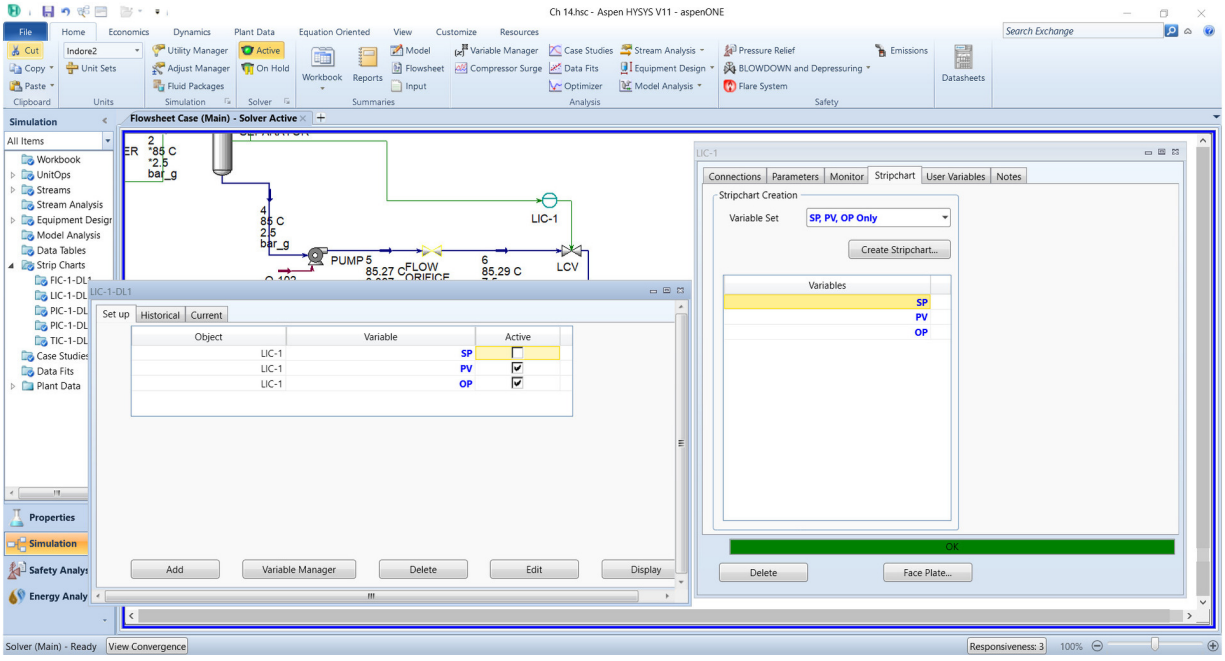
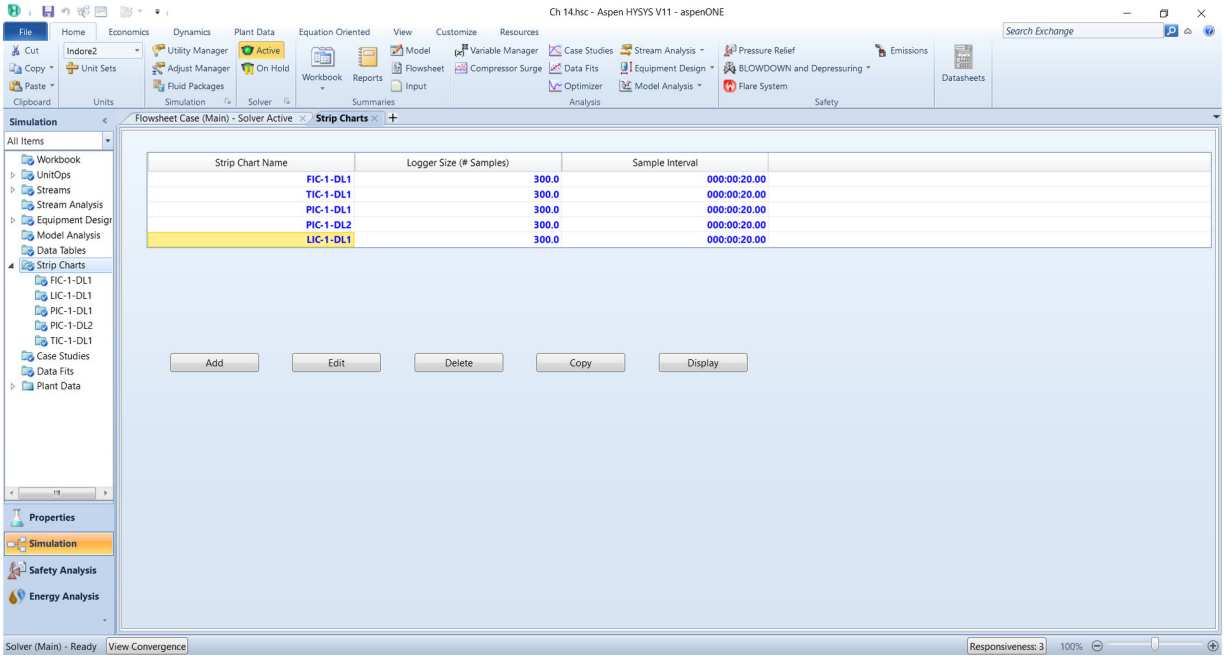
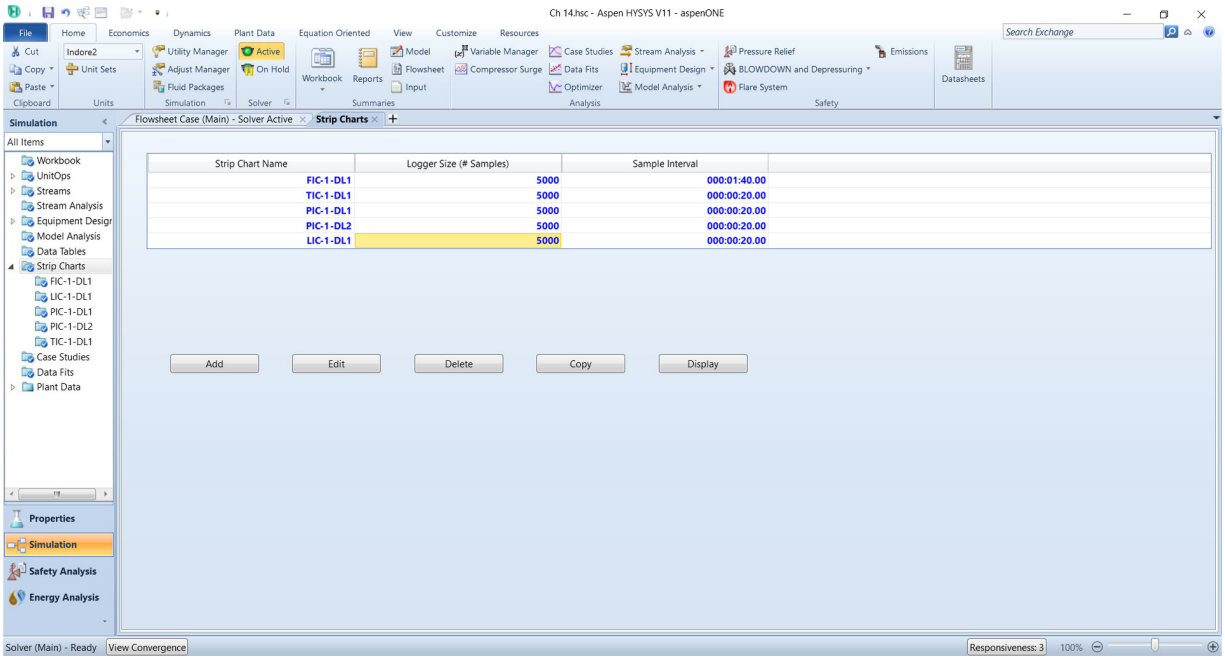


Figure 16.46: Strip chart



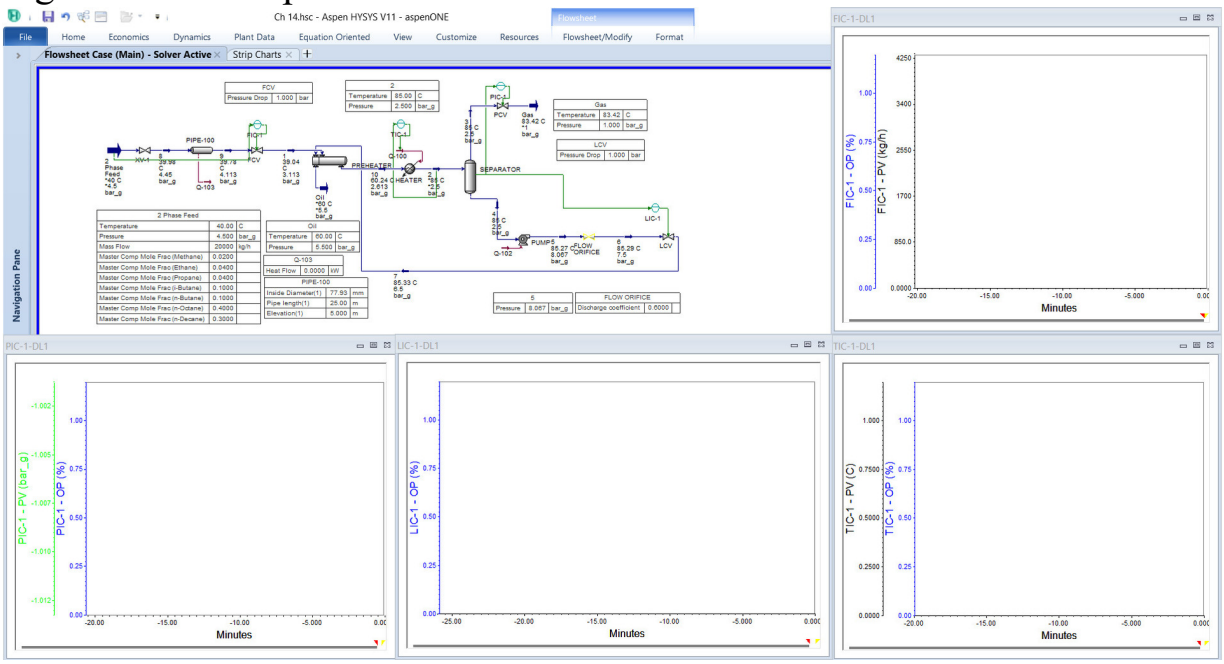
Increase the “Logger Size” to 5000.

Figure 16.47: Strip chart



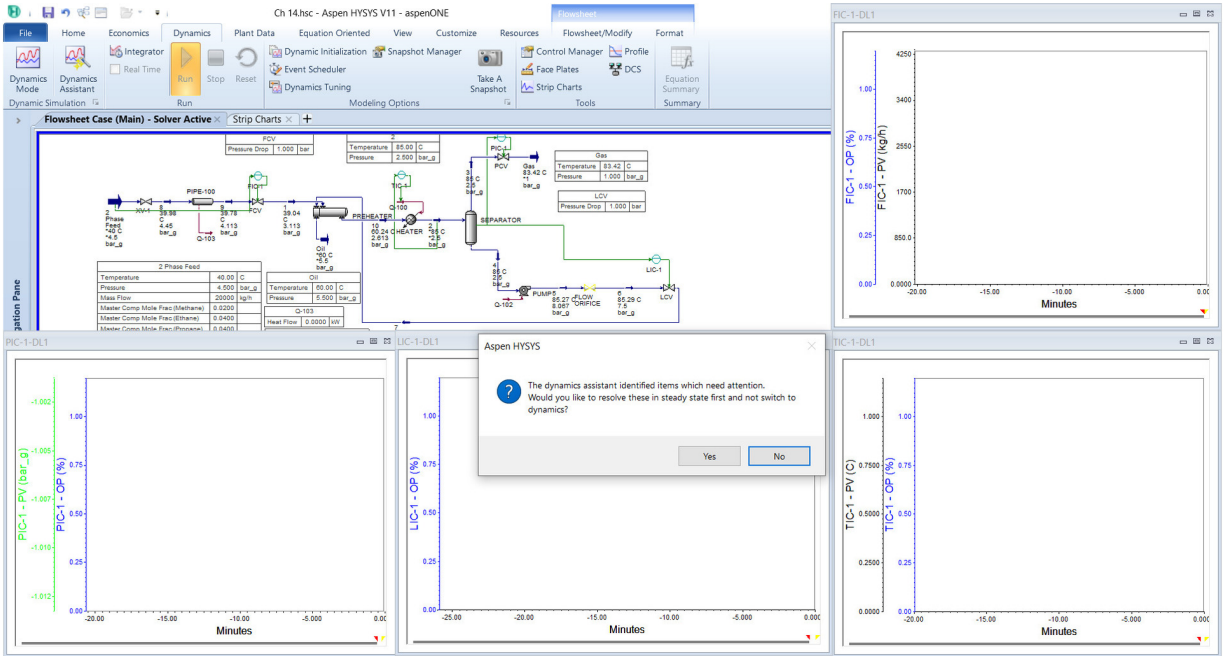
Arrange all the 4 plots and the model as shown below.

Figure 16.48: Strip chart



Now we are ready to convert from steady-state to dynamic mode. Click on "Dynamics Mode". If there are some issues in the model that needs attention, then it shall be pointed out by "Dynamics Assistant".

Figure 16.49: Strip chart



The suggested remedies will be listed by “Dynamics Assistant”. One can check each item and make changes, if needed, or click directly on “Make Changes” so that “Dynamics Assistant” makes changes for you in the model.

Figure 16.50: Dynamic assistant

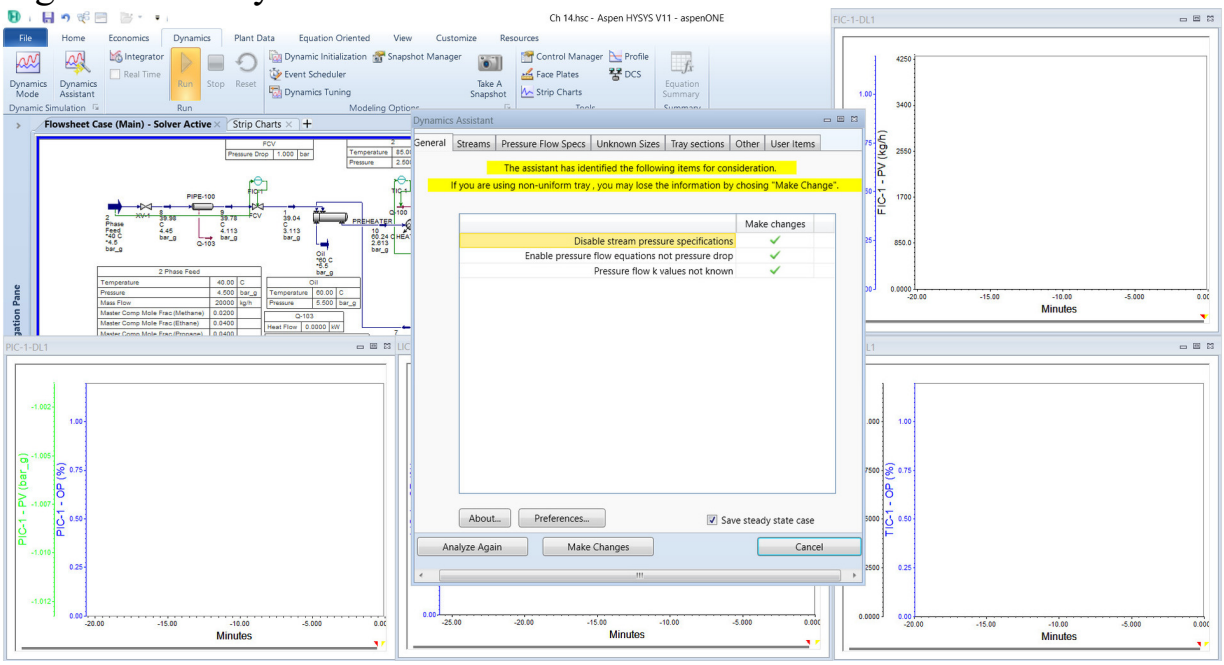
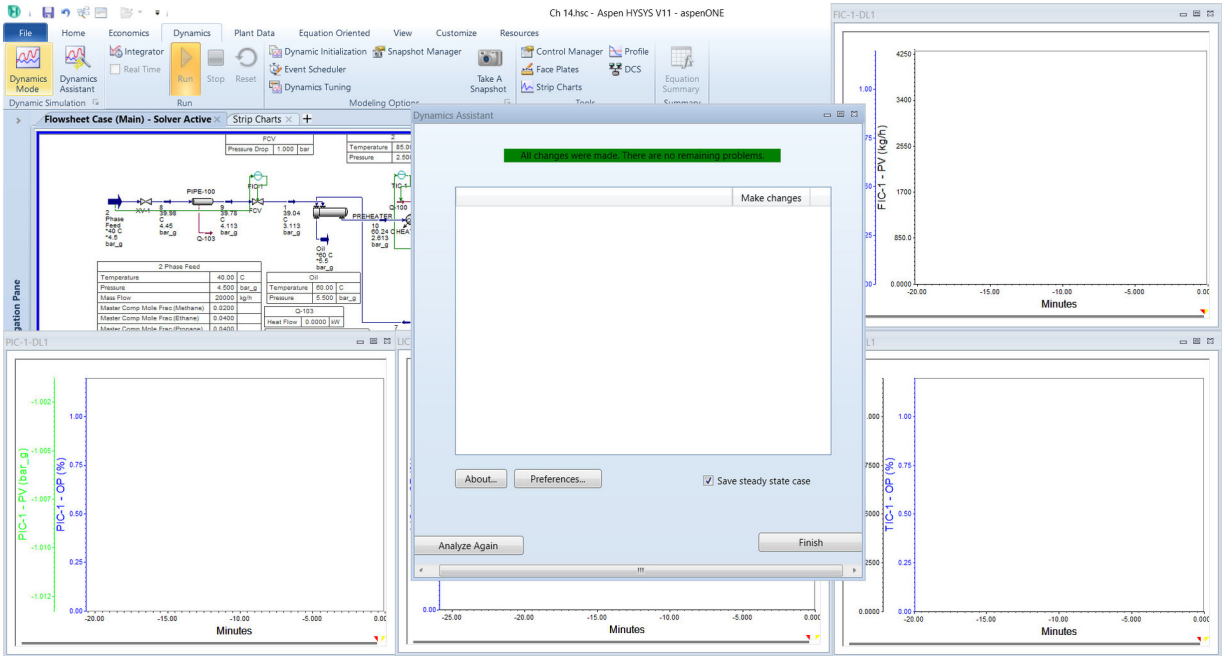
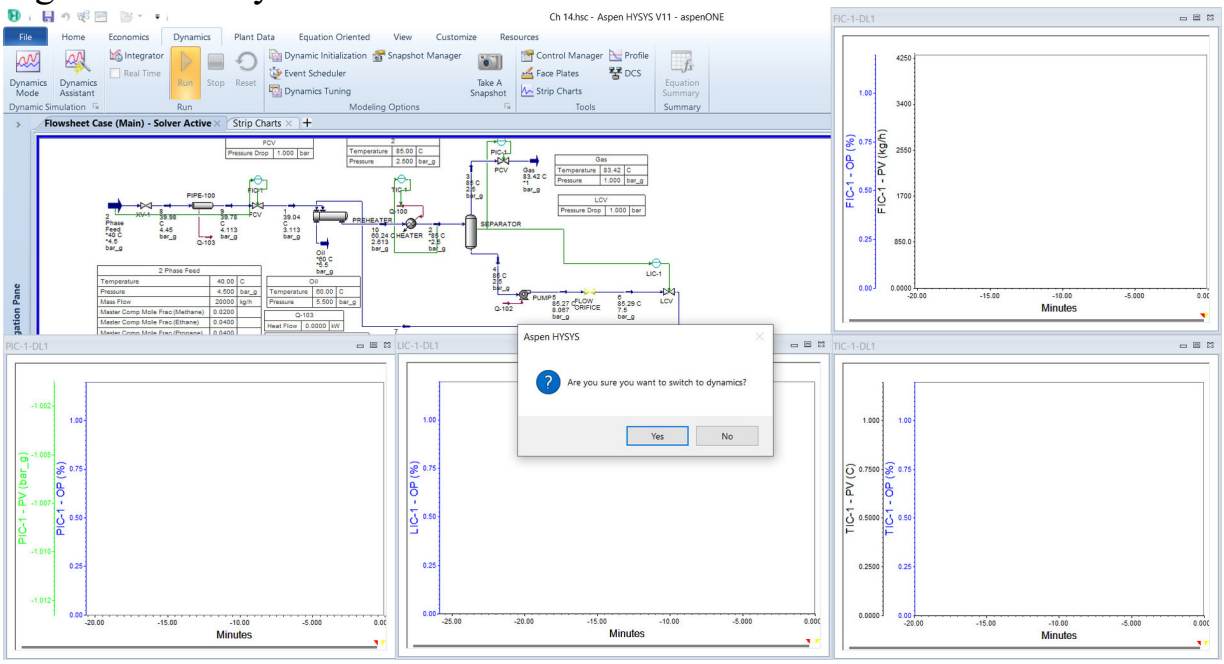


Figure 16.51: Dynamic assistant



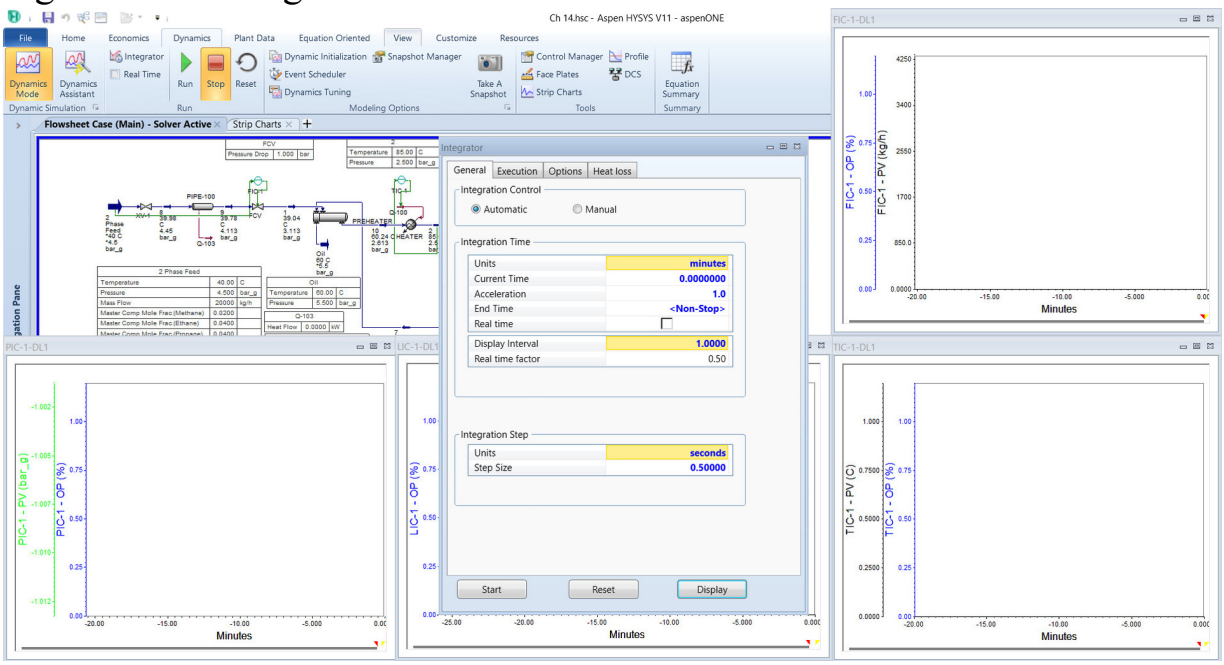
Click “Dynamics Mode” to enter Dynamics mode.

Figure 16.52: Dynamic mode



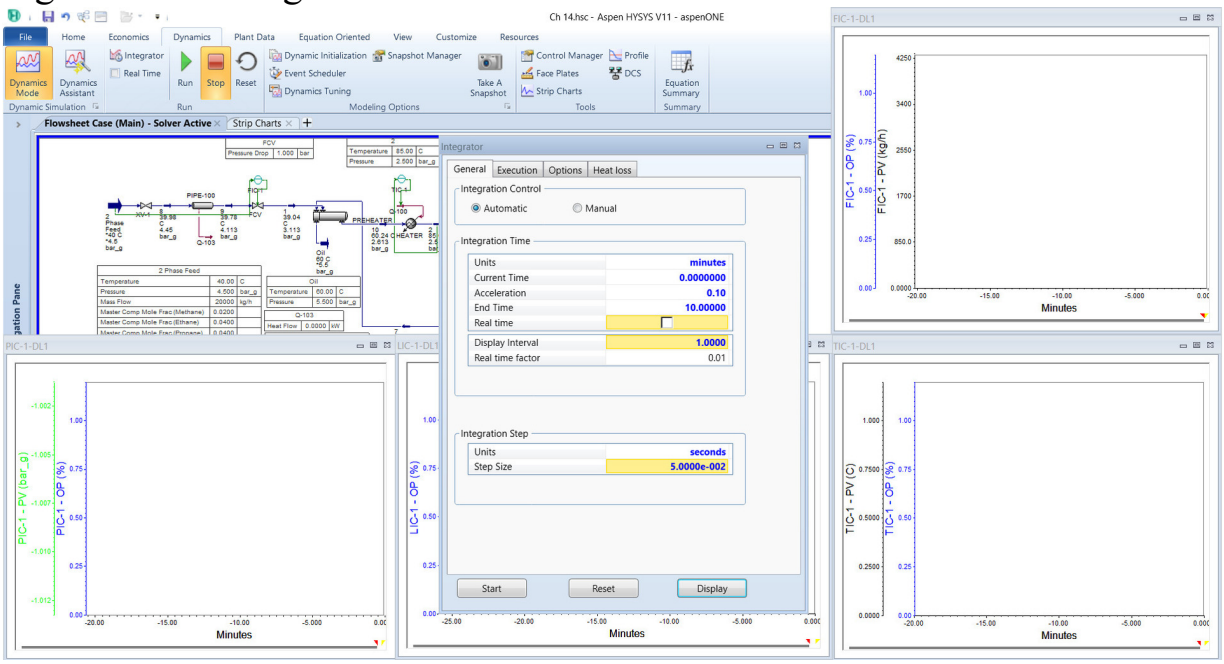
Click on “Integrator”.

Figure 16.53: Integrator



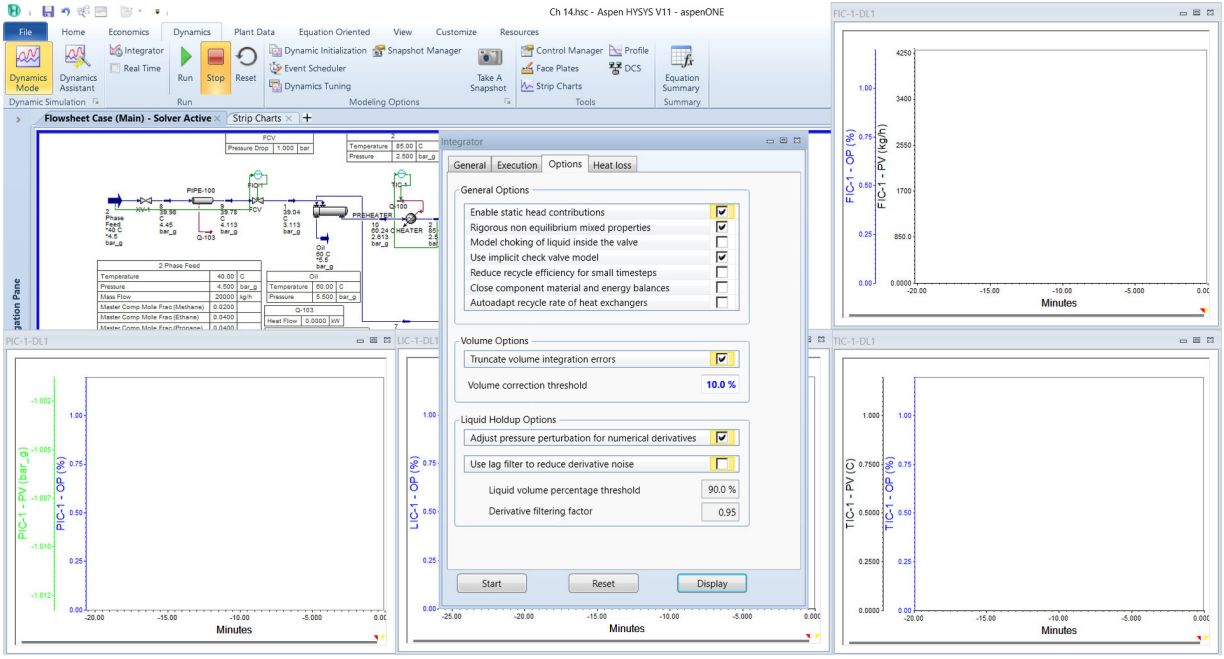
We can assign run duration, speed, and Calculation step size in “Integrator”.

Figure 16.54: Integrator



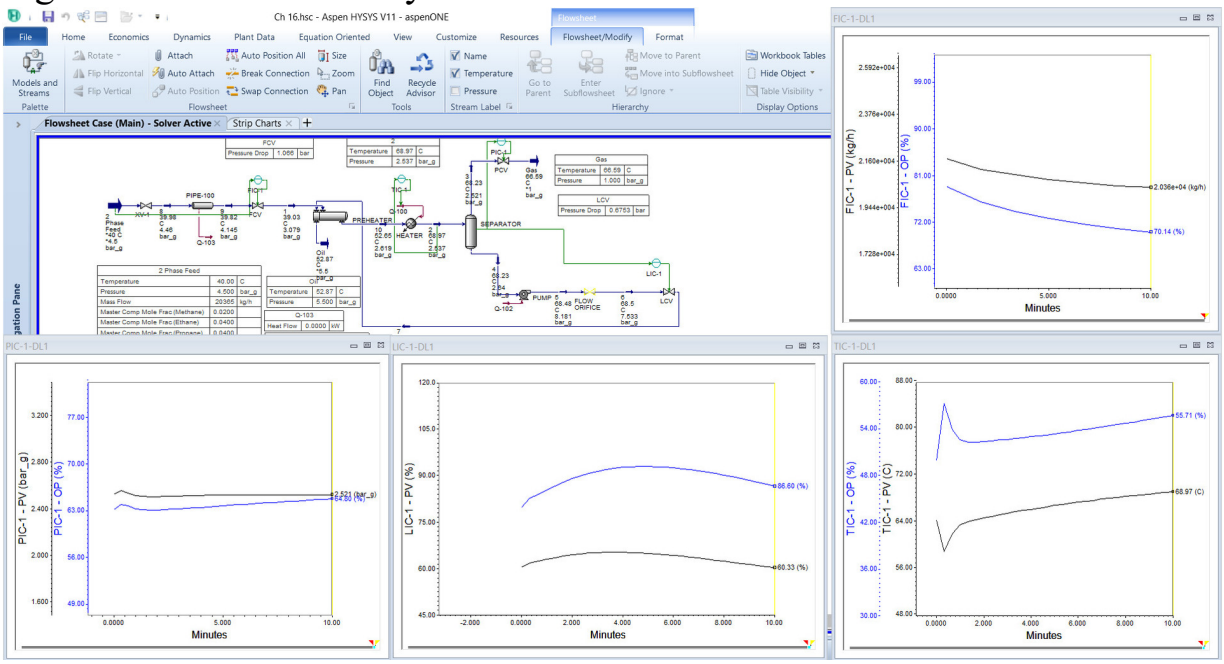
Check “Enable static head contributions” in the “Integrator\Options\General Options” tab.

Figure 16.55: Integrator



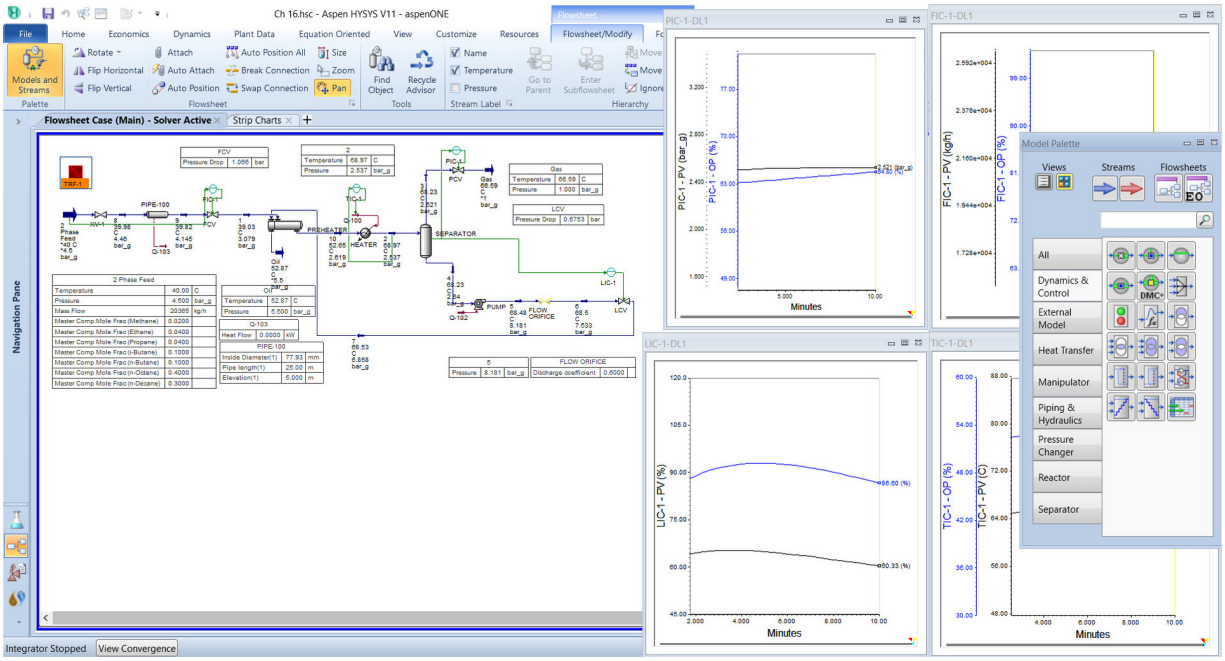
Run the model for 10 minutes and analyze the results.

Figure 16.56: Model analysis



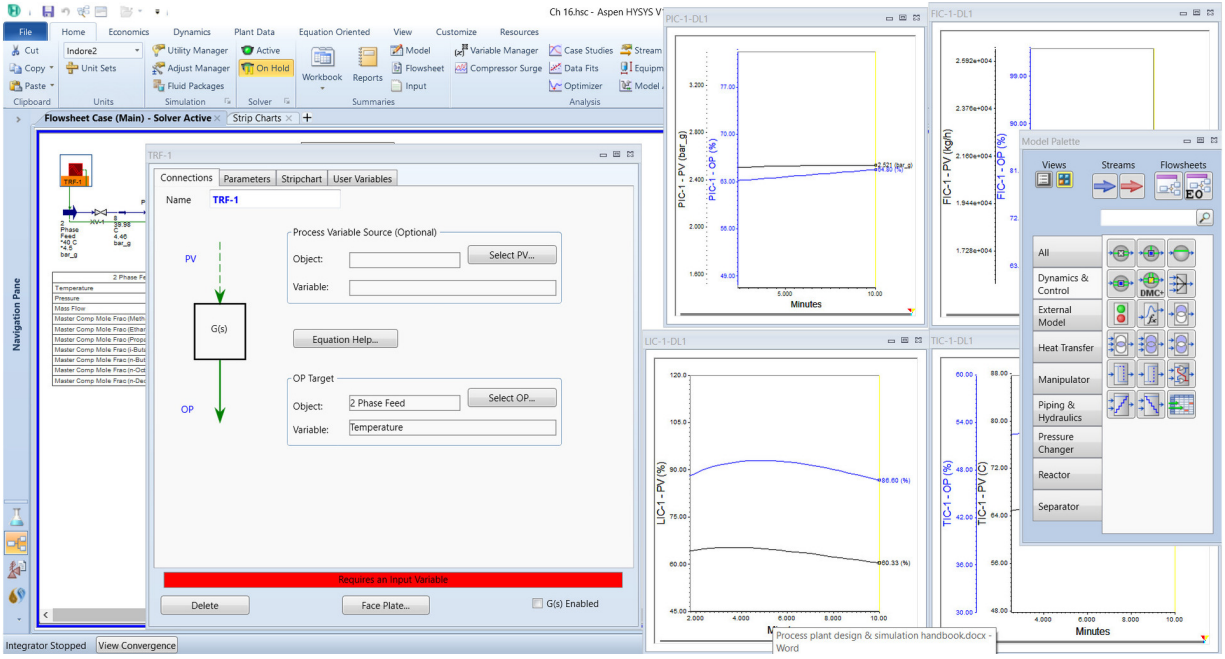
The Transfer Function block is a logical operation that takes a specified input, and applies the chosen transfer function to produce an output. A typical use of the Transfer Function is to apply disturbances to a process, such as varying the temperature of a feed stream.
Insert “Transfer Function Block” from the model palette.

Figure 16.57: Transfer function block



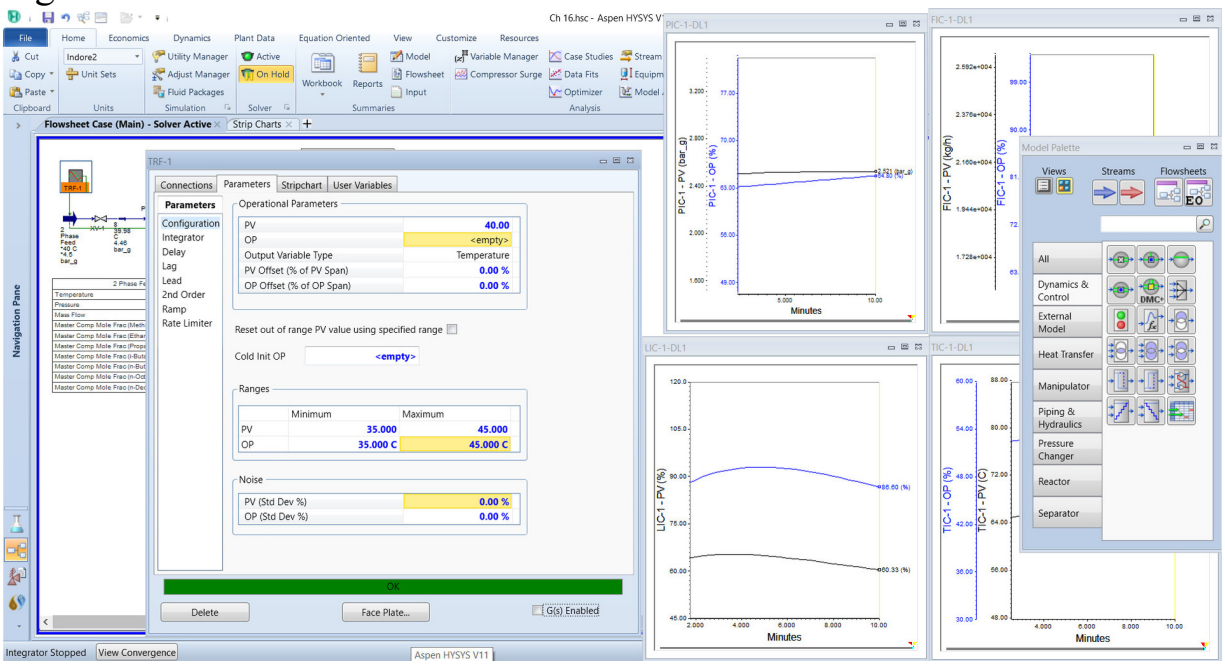
We will keep changing 2 Phase feed stream temperature between 20 ° C to 60 ° C using “Transfer Function Block” (TRF-1).

Figure 16.58: Transfer function block



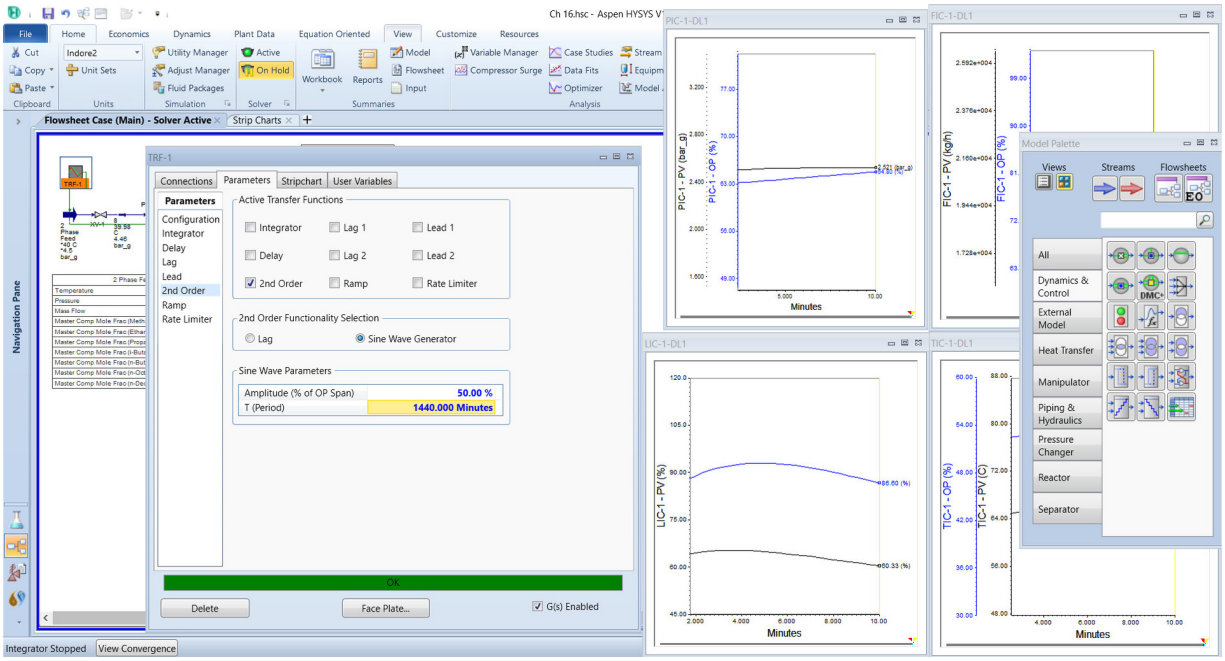
“PV” is a process variable (and the present value is 40 ° C). PV and OP ranges will be kept as 20 ° C to 60 ° C.

Figure 16.59: Transfer function block



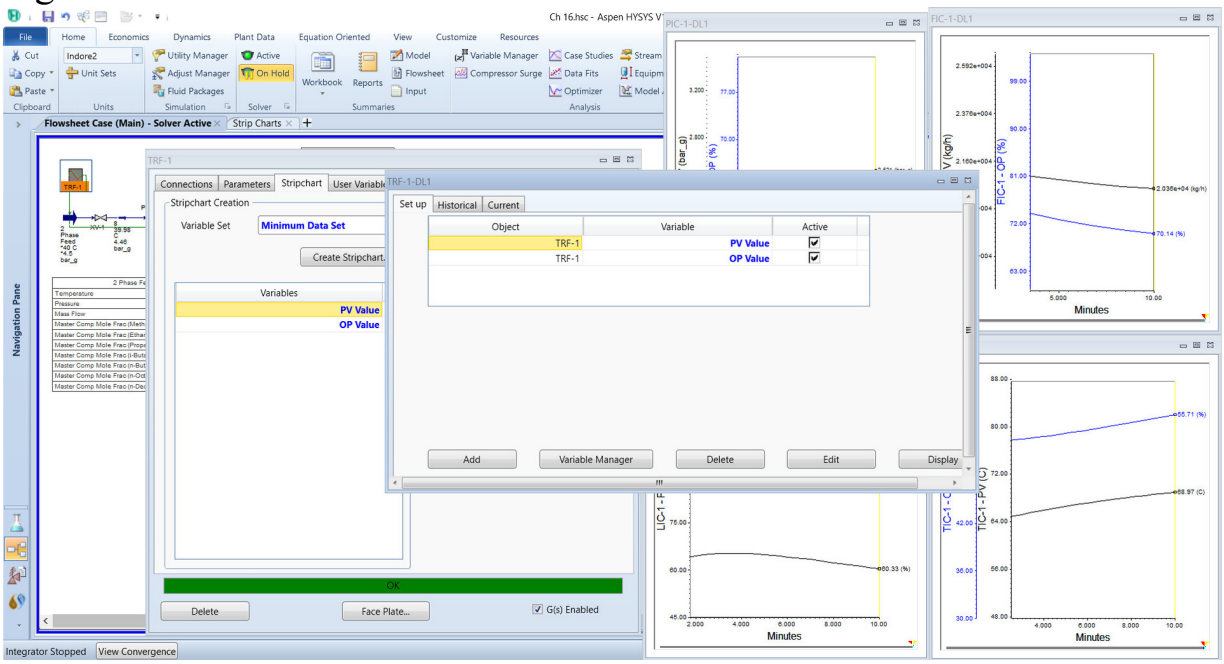
OP span is (45 – 35 =) 10 ° C. Amplitude (% of OP span) is 50% or 5 ° C. The total period is considered as 24 hours (1440 minutes) in the simulation, meaning in 24 hours duration stream temperature will reach as high as 45 ° C or as low as 35 ° C. “Active Transfer Functions” selected is “2nd Order” and “2nd order Functionality Selection” is “Sine Wave Generator”.

Figure 16.60: Transfer function block


Ch 16.hsc - Aspen HYSYS V
File
Home
Economics
Dynamics
Plant Data
Equation Oriented
View
Customize
Resources
Flowsheet Case (Main) - Solver Active
Strip Charts
Navigation Pane
TRF-1
Connections
Parameters
Stripchart
User Variables
Configuration
Integrator
Delay
Lag
Lead
2nd Order
Ramp
Rate Limiter
Active Transfer Functions
Integrator
Lag 1
Lead 1
Delay
Lag 2
Lead 2
2nd Order
Ramp
Rate Limiter
2nd Order Functionality Selection
Lag
Sine Wave Generator
Sine Wave Parameters
Amplitude (% of OP Span)
50.00 %
T (Period)
1440.000 Minutes
OK
Delete
Face Plate...
G(s) Enabled
Integrator Stopped
View Convergence
PIC-1-DL1
PIC-1 - PV (barg)
PIC-1 - OP (%)
Minutes
FIC-1-DL1
FIC-1 - PV (kg/h)
FIC-1 - OP (%)
Model Palette
Views
Streams
Flowsheets
All
Dynamics & Control
External Model
Heat Transfer
Manipulator
Piping & Hydraulics
Pressure Changer
Reactor
Separator
LIC-1-DL1
LIC-1 - PV (%)
TIC-1-DL1
TIC-1 - OP (%)
TIC-1 - PV (C)

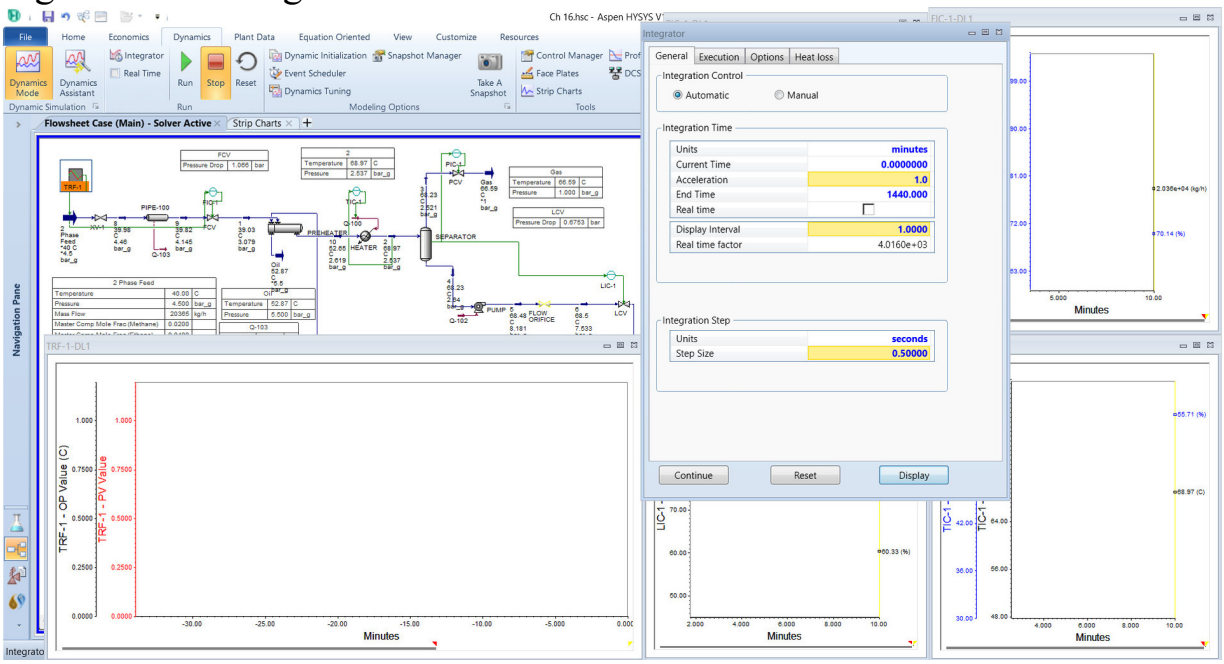
Provide a strip chart for “TRF-1” as well. The sample size for this strip chart will be the same as the other 4 plots.

Figure 16.61: Transfer function block



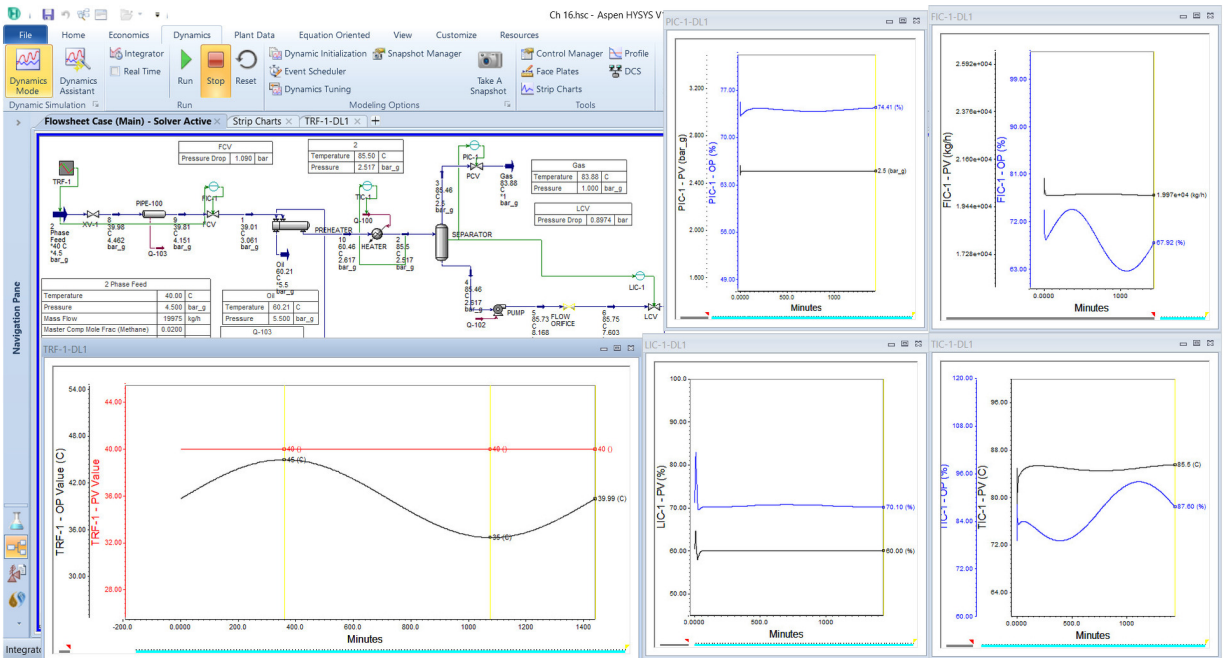
Open “Integrator” and make changes as below.

Figure 16.62: Integrator



The temperature fluctuation is 24 hours for 2 Phase Feed stream is given in the below plot. And also, the impact of the same on the other 4 controllers are provided.

Figure 16.63: Model analysis



CHAPTER 17: TYPICAL INTERVIEW QUESTIONS

The following are the most commonly asked technical questions for a process engineering interview. The reader is expected to write down all the answers for the same.

1. Please introduce yourself.
2. What is your present job profile?
3. Why do you want to switch jobs?
4. Why do you think this company (where you have applied) is better suited for your career goals?
5. What is your favorite topic in chemical engineering? Why?
6. Have you designed a process plant before?
7. How good are you with MS excel?
8. How to carry out heat exchanger sizing?
9. What do you mean by pump hydraulic power?
10. What is $NPSH_{available}$?
11. Explain control valve sizing procedure.
12. What is cavitation in control valves?
13. What are the different standards used in the process industry?
14. What are the different types of PSVs?
15. Explain Table 1 of API 520 Part 1.
16. Explain different line sizing criteria.
17. What is the L/D ratio of horizontal separator?
18. What are the different selection criteria for separators?
19. What are the different constraints for orifice sizing?
20. What do you mean by valve Cv?
21. What is meant by Critical Pressure and Critical Temperature?
22. Why is reduced trim required in control valves?

23. What are the different line sizing criteria?
24. Explain pump sizing steps with an example.
25. Explain control valve sizing steps with an example.
26. What are the control valve flow characteristics?
27. What is temperature cross in the heat exchanger?
28. What are the different process standards used in the industry?
29. Explain pressure safety valve sizing steps with an example.
30. Why is typically TRV size D?
31. Explain orifice sizing steps with an example.
32. What is the compressor settle out pressure?
33. What do you mean by compressor surge?
34. Explain blow down criteria.
35. Explain the functioning of the cooling tower.
36. What are the different types of agitators?
37. Explain heat exchanger sizing steps with an example.
38. How is the reactor size decided?
39. Explain Heat Exchanger control and Interlock and Control Philosophy related to Fired Heater? (Toyo Engineering)
40. Explain Feedback and Cascade Control? Draw Feedback and Cascade Control Loop? (Toyo Engineering).
41. Why Line Sizing is important? How Line Sizing will be carried out? (Toyo Engineering)
42. Velocity guidelines for Liquid, Gas, Steam and vacuum applications? (Toyo Engineering)
43. What is Schedule no. in Line Size? What significance it has in line sizing? (Toyo

Engineering)

44. What factors affect pressure drop? (Toyo Engineering)
45. How to calculate Heat Duty of Heat Exchanger? (Toyo Engineering)
46. Show P&ID representation of the vessel where the control valve is controlling the level of the vessel? (Toyo Engineering)
47. What is PSV? Criteria for sizing Inlet line of PSV? (Toyo Engineering)
48. Criteria for sizing tailpipe of PSV and Flare Header? (Toyo Engineering)
49. If Backpressure is <10%, which PSV shall be used? (Toyo Engineering)
50. Draw Centrifugal and Positive Displacement Head, Power, and Efficiency curves? (Toyo Engineering)
51. What is the best efficiency point of the pump? (Toyo Engineering)
52. Explain Heat Exchanger control and Interlock and Control Philosophy related to Fired Heater? (Toyo Engineering)
53. Explain in Detail Control Philosophy and Interlock Philosophy of Fired Heater? Draw entire P&ID for the same? (Toyo Engineering)
54. Explain IBR PSV? Which Code to be applied for PSV related to IBR Pressure Vessel? (Toyo Engineering)
55. What are the criteria to size PSV? (Toyo Engineering)
56. What is the Overpressure? (Toyo Engineering)
57. Why Set pressure of PSVs are kept different? (Toyo Engineering)

58. Draw P&ID of Fired Heater in detail and explain each Interlock and explain control philosophy in brief? (Jacobs India)
59. What role does DCS and PLC play? Which are the symbols and how to represent in P&ID? (Jacobs India)
60. How do you calculate Heat Duty of Heat Exchanger? (Jacobs India)
61. What is NPSH required and cavitation? (Jacobs India)
62. How to prevent cavitation in pumps? (Jacobs India)
63. How do you calculate NPSH? (Jacobs India)
64. How to calculate the power of a pump? (Jacobs India)
65. How do you calculate Line size and pressure drop? (Jacobs India)
66. Which formula is used to calculate Pressure drop? (Jacobs India)
67. How friction factor will affect pressure drop? (Jacobs India)
68. What is roughness and does it change with the MOC of pipe? If roughness is more, will pipe pressure drop be more? (Jacobs India)
69. What is PSV? What is the overpressure considered for the Fire case scenario? (Jacobs India)
70. What are the different types of PSV? (Linde India)
71. Blocked case or Fire case – which PSV is bigger? (Linde India)
72. What are the different codes used in PSV? (Linde India)

73. What is the pump shut off pressure? (Linde India)
74. What is the formula for shut off pressure? (Linde India)
75. What is depressurizing utility and the rule for depressurizing? (Linde India)
76. What are the line size criteria for PSV tailpipe & flare header? (Linde India)
77. What is the line sizing criteria for the blowdown line? (Linde India)
78. What are the line sizing criteria for pump suction & discharge line? (KBR India)
79. Discuss instrumentation and valves for area around centrifugal pump. (KBR India)
80. Discuss Hazen Williams formula? (KBR India)
81. Is PSV set pressure and pop up pressure the same? (KBR India)
82. Explain the refrigeration cycle. (KBR India)
83. What is the maximum velocity for the vapor line before and after the compressor? (KBR India)
84. What are the effects of the pump operating away from BEP?
85. What is Bernoulli's principle?
86. What is meant by the pipe schedule?
87. What is meant by the pump shut off head?
88. Explain the Froude number.
89. Write Mach number ranges for different types of flow.
90. Is control valve is an on-off valve? (LoneStar Technology)
91. What is a check valve? (LoneStar Technology)
92. What are the different types of valves? (LoneStar Technology)

93. What are the input parameters inline sizing? (LoneStar Technology)
94. How Orifice will work? (LoneStar Technology)
95. Where the pressure levels are high at the Orifice Upstream or the downstream? (LoneStar Technology)
96. What are the input parameters in pump sizing? (LoneStar Technology)
97. Why suction pressure has to be higher than the flowing liquid vapor pressure? (LoneStar Technology)
98. What is cavitation? (LoneStar Technology)
99. How can cavitation be eliminated? (LoneStar Technology)
100. How many types of flow are there in the Heat exchanger? (LoneStar Technology)
101. What is energy balance? (LoneStar Technology)
102. On what principle distillation column work? (LoneStar Technology)
103. What are the different outputs of the crude distillation? (LoneStar Technology)
104. What is chromatography? (LoneStar Technology)
105. What is PFD? (LoneStar Technology)
106. What are the differences between PFD and P&ID? (LoneStar Technology)
107. What are the instruments you can provide in the P&ID? (LoneStar Technology)
108. What is the purpose of the cause and effect matrix? (LoneStar Technology)
109. What is FEED? (LoneStar Technology)
110. What is the role of process engineer?

111. What are the various unit operations used in process industries?
112. What is the purpose of PFD?
113. List out the differences between PFD and P&ID.
114. What is the difference between Cv and Kv?
115. Define compressible flow and incompressible flow?
116. Define choked flow and non-choked flow?
117. What is Reynolds Number?
118. What are the input parameters required to size the control valve?
119. Describe pressure drop criteria in the selection of Control valve?
120. What is the objective of pressure safety valve sizing?
121. Define set pressure and accumulation.
122. What are the set pressure and Accumulation limits for pressure relief valves?
123. Differentiate between superimposed backpressure and built-up back pressure?
124. What is meant by chattering? How can it be eliminated?
125. What are the causes of chattering?
126. Write about Chattering solutions?
127. Is design pressure and MAWP same?
128. Differentiate between accumulation and blowdown?
129. Is it good to use a single PSV for two equipment/vessels? Why?
130. In which cases one selects multiple PSV?
131. What is the role of MAWP to design PSV?
132. What do you understand by the term depressuring rate?

133. What are the input parameters required to design a PSV?
134. What are the guidelines to follow in the line sizing?
135. Is the pump suction line greater than the discharge line? Why?
136. What is meant by erosion velocity?
137. What is the objective of line sizing?
138. What is Stokes law?
139. What is the friction factor?
140. What are STP conditions and NTP conditions?
141. What are the different types of pumps used in the industry?
142. Explain the necessity of the performance curves to size the pump.
143. Differentiate between pump and compressor.
144. Is total developed head and differential head the same? What the units?
145. Define specific gravity.
146. How to calculate $NPSH_{available}$?
147. Should $NPSH_{available}$ is greater than $NPSH_{required}$? Why?
148. Explain the effect of Specific Gravity to calculate Head?
149. Explain VLE?
150. What is relative volatility?
151. How many types of separators are available?
152. How to design heat exchanger design using Aspen EDR?
153. What is TEMA sheet?
154. When do we use heat exchangers in series or parallel?
155. How to increase the turbulence on the shell side and tube side?

156. What is temperature cross in the heat exchanger?
157. What is meant by LMTD?
158. What is the role of baffles in the heat exchanger?
159. What is fouling and its units?
160. How many types of baffles are available?
161. On what principle distillation column works?
162. What is Raoult's law?
163. Describe the use of the McCabe Thiele method.
164. What is the effect on the distillation column load if the reflux ratio is increased?
165. Define diffusivity.
166. What are the units of molar flux?
167. What is superficial velocity?
168. For pure liquid, what is bubble point temperature and dew point temperature?
169. What is relative volatility?
170. What is minimum boiling azeotrope and maximum boiling azeotrope?
171. What are the different types of distillation?
172. Provide differences between partial reboiler and total reboiler.
173. Provide differences between partial condenser and total condenser.
174. What is meant by Azeotrope?
175. What are the advantages of distillation?
176. What are the limitations of the distillation column?
177. Define vapor pressure and Reid vapor pressure.
178. Define Sensible heat and latent heat.
179. What is the purpose of the flare header?
180. What are flare system components?

181. Write about aspen Flare System Analyzer.
182. What is a flare stack?
183. Differentiate between BDV and PSV.
184. What are the design criteria for BDV?
185. Is it necessary to provide BDV and why?
186. To which system is the BDV downstream connected?
187. What is the purpose of an orifice plate?
188. What is the difference between flow orifice and restriction orifice?
189. What is the best location for flow orifice, upstream, or downstream of the control valve? Why?
190. On what principle Orifice works?
191. What is heat exchanger fouling? Using equations, please explain how heat exchanger fouling impacts the duty of heat exchangers.
192. Using equations, explain the impact of the viscosity of the heat transfer fluid on heat exchanger duty.
193. In a pipe, consider that there is a turbulent flow of fluid whose composition is known. If you know the diameter of the pipe, as well as the length of the pipe, can you calculate the flowrate of through the pipe, if you know the pressure at the inlet and outlet of the pipe? Explain your answer.
194. Consider a compressor that is designed for a certain gas mixture. What would happen if the gas mixture is changed?
195. In a distillation column, consider that we have a mixture of chemicals being fed. One of the components is in a small amount. As you perform a distillation column simulation, you

find that the component is going in the bottoms. However, your friend, who is working in the plant which operates this column says that this component is going to the top of the column, what would you do to modify the simulation.

196. In a distillation column, you have a mixture of benzene, toluene, and all the xylene isomers. If you want to separate benzene and toluene at the top, and xylenes at the bottom, can you estimate the temperature which would be at the top of the column and the temperature at the bottom of the column? Explain your answer.

197. In a distillation column, you are feeding a mixture of hydrocarbons with a vapor mass fraction of 0.7 and 10000 m^3 /h close to the bottom of the column. If the column has no reboiler, but the column bottom is fed with a small amount of live steam, can you estimate the liquid flow at the bottom of the column? You can make some assumptions about the steam rate.

198. In the distillation column above, there is a condenser at the top of the column, but no other side coolers / pump-around. The top condenser condenses all vapor, and part of the condensed vapor is used as reflux, and part is drawn as a top product. What will the top distillate rate? You can make some assumptions about the reflux rate.

199. In the case of a reactor, we find that a set of components A, B, and C react, and form components D and E. If all the reactions are very fast, what approach would you take to

estimate the composition in the reactor product?

200. In a combustion reactor, a mixture of methane, ethane, and air is fed at roughly 50 ^{o} C. Due to the combustion reactions, the temperature at the outlet of the reactor will rise. How much air will be required if an equimolar mixture of methane and ethane at 1000 Nm^3 /h are reacted? What will be the outlet temperature? Explain your approach to solving this problem.
201. In the above example, if we decide to use enriched oxygen instead of air, what will it do, and why?
202. In a process, we are reacting a liquid hydrocarbon mixture with hydrogen gas in the gas-liquid mixed reactor. The reactor is followed by a separator, and the leftover gas is separated, compressed and recycled back, while the liquid leaves the system. Do you need a purge gas stream? Explain your answer. Do you need a makeup gas stream? How will you calculate the makeup gas flow? (Jacobs India)
203. What is the effect of flowrate on the head? (Jacobs India)
204. Draw Pump characteristics curve. (Jacobs India)
205. Explain different types of PSV? How does it work? (Jacobs India)
206. Describe the pressure drop criteria of the control valve. (Jacobs India)
207. Is PSV set pressure and design pressure the same or not? (Jacobs India)

208. How to carry out the PSV inlet line? (Jacobs India)
209. What information does the line list contain? (TechnipFMC India)
210. What information does the equipment list contain? (TechnipFMC India)
211. What does P&ID contain? (TechnipFMC India)
212. Why is PSV used? (TechnipFMC India)
213. What are the different types of PSVs? (TechnipFMC India)
214. How to control surge? (TechnipFMC India)
215. Provide $NPSH_A$ formula (Saipem India)
216. Which should be higher $NPSH_A$ or $NPSH_R$? (Saipem India)
217. What is shutoff pressure? (Saipem India)
218. Mark shutoff pressure in the graph. (Saipem India)
219. Draw P&ID for storage tank with pump at downstream and mention the instrument needed. (Saipem India)
220. List of documents required for P&ID making. (Samsung Engineering)
221. What are different safety valve sizing scenarios? (Samsung Engineering)
222. What is meant by NPSH of the pump? (Samsung Engineering)
223. What is the output of pump hydraulics study? (Samsung Engineering)
224. Describe the principle of blowdown. (Samsung Engineering)
225. Explain pressure drop criteria for line sizing. (Samsung Engineering)
226. Describe gas sweetening process. (Samsung Engineering)

227. Explain the procedure for control valve sizing. (Bechtel India)
228. Explain the effect of higher velocities in the pipeline. (Bechtel India)
229. Describe the PFD of the project you worked on. (Bechtel India)
230. Explain different line sizing criteria. (Bechtel India)
231. How one can reduce the pressure drop across the control valve? (Bechtel India)
232. What is the formula for erosional velocity? (Bechtel India)
233. What are pump curves? (Ingenero India)
234. How to carry out line sizing and pressure drop calculation. (Ingenero India)
235. Explain any process by using a block diagram. (Ingenero India)
236. Explain the distillation column control system. (Ingenero India)
237. What are the causes of Overpressure in the column? (Ingenero India)
238. What are the pump overpressure causes? (Ingenero India)
239. Explain every unit of the refinery in with its significance.
240. What are the simulation steps for distillation column sizing?
241. Explain TEMA and the significance of its parameters.
242. What is the significance of ρv^2 in the heat exchanger?
243. What is better; crossflow or counter-current flow in a heat exchanger?